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Prepared for

Industrial Environmental Research Lab, Research Triangle Park, NC

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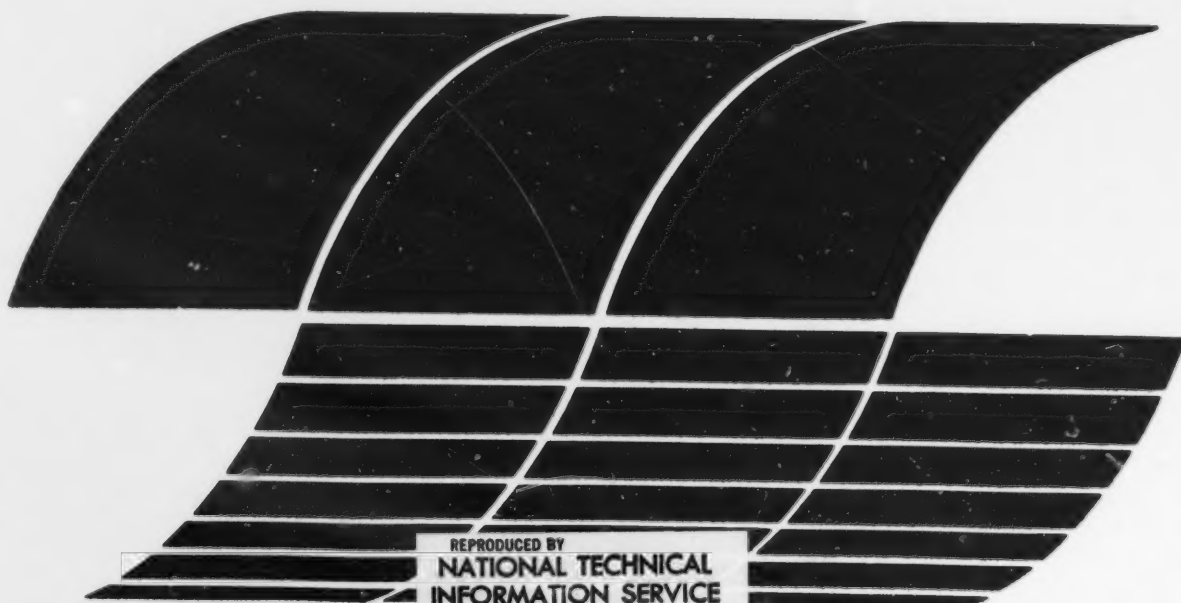
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**Interagency
Energy/Environment
R&D Program Report**



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Franklin A. Ayer, Compiler

Research Triangle Institute
P. O. Box 12194
Research Triangle Park, North Carolina 27709

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Industrial Environmental Research Laboratory
Office of Energy, Minerals, and Industry
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ABSTRACT

This publication contains the text of all papers presented at EPA's 5th FGD Symposium held in Las Vegas, Nevada on March 5-8, 1979. Papers cover such subjects as health effects of sulfur oxides, impact of FGD on the economy and the energy problem, energy and economics of FGD processes, actual operating experience, waste disposal and byproduct marketing, and industrial boiler applications.

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BASIN ELECTRIC'S INVOLVEMENT
WITH DRY FLUE GAS DESULFURIZATION

Kent E. Janssen
Manager of Production
and
Robert L. Eriksen
Environmental Control Supervisor

Basin Electric Power Cooperative
1717 East Interstate Avenue
Bismarck, ND 58501

ABSTRACT

This report will introduce a relatively new technique of flue gas desulfurization using the dry scrubbing process. The dry scrubbing system introduces a reagent slurry, via an atomizer into the flue gas where the SO₂ is absorbed by the reagent particles. The heat from the flue gas dries the particles in the reaction chamber. The particles are then collected in an electrostatic precipitator or a baghouse and disposed of dry, eliminating the sludge dewatering process needed for the wet scrubbing systems. Several reagents have been tested with soda ash and lime yielding the most favorable results. Atomization can be accomplished by either a rotary atomizer or spray nozzle. Baghouses and precipitators are known equipment in utility plants and spray dryers are operated in thousands of chemical and industrial plants around the world.

Basin Electric Power Cooperative will use the rotary atomizer and baghouse collection at the Antelope Valley Station Unit 1 (440 MW) near Beulah, ND. The "Y" jet spray nozzle and precipitator collection will be used on the third unit (500 MW) of the Laramie River Station near Wheatland, WY. Reagents for both systems will be lime, due to its favorable economics over soda ash. It is estimated that the dry scrubbing system will save approximately \$23 million at the Laramie River Station and \$47 million at the Antelope Valley Station over the life of the plants compared to a "wet scrubbing" system for these site specific installations.

We believe the utility industry will eventually accept the concept of dry scrubbing, not so much due to the economic factor, but because of the system's simplicity and potential high availability.

BASIN ELECTRIC'S INVOLVEMENT WITH DRY FLUE GAS DESULFURIZATION

INTRODUCTION

Numerous processes are available for removing SO_2 from the flue gas produced in fossil-fired power plants. Lime or limestone "wet scrubbing" processes are the most common used today. The wet scrubbing systems have generally been promoted by government agencies but have found few proponents among the electric utility industry. The wet scrubbers are complex and cumbersome. Added to this, wet scrubbers cause a heavy capital burden, have a historic low availability record, are costly to operate, difficult to maintain and consume a sizeable part of the energy generated by the plants they serve. The byproduct, sludge, produces additional environmental problems by taking up otherwise useful land for settling ponds.

We question the rationale which dictates the current and proposed air quality standards requiring high removal rates of SO_2 , particularly for the low sulfur coals. We question that sufficient research has been done to determine the need for such strict standards and whether the regulations will improve our air quality sufficiently to justify the high cost of removal. Yet, in order to meet the standards in a manner which is more operationally acceptable, we have worked with several manufacturers to develop a process and hardware which is less complex and, for our situation, more economical as well.

We believe the development of "dry scrubbers" during the past two years may be a major improvement in the field of flue gas desulfurization.

HISTORY OF DRY FLUE GAS DESULFURIZATION

Basin Electric's involvement with dry scrubbing began with a pilot baghouse at our Leland Olds Station near Stanton, North Dakota, in 1976. The program was developed for the Coyote Project, a consortium of five utilities headed by Otter Tail Power Company. Unit 2 of the Leland Olds Station was selected because it is a 440 MW cyclone fired boiler which burns North Dakota lignite, therefore having conditions very similar to the proposed Coyote Station. Bechtel Power Corporation, the architect-engineer for Coyote, coordinated the program. The Wheelabrator-Frye, Inc. (WFI) baghouse used a process of injecting dry powdered nahcolite into the flue gas stream and onto fabric filter bags to remove SO_2 . This process worked quite well, but problems developed in the availability of nahcolite. Major nahcolite reserves are in Colorado, but federal regulations make mining of it very difficult. As a result, the long range outlook for nahcolite was not good because of supply uncertainties. This led to the investigation of other processes. In the spring of 1977, preliminary testing began on the open loop portion of the Rockwell International, Atomics International Division Aqueous Carbonate Process, utilizing a Bowen spray dryer with a rotary disc atomizer, followed by a WFI baghouse.

The first reagents used at the pilot plant were soda ash and trona. These produced excellent SO₂ removals and reagent utilization in the dry scrubber.

PILOT PLANT TESTING

We, at Basin Electric, were interested in the pilot work as we were involved in a large construction program with all units requiring scrubbers-- the Antelope Valley Station (two 440 MW units) near Beulah, North Dakota and the Laramie River Station (three 500 MW units) near Wheatland, Wyoming; with plans for future power plants. Because of the potential disposal problems and the high cost of sodium reagents, we decided to investigate the use of other alkaline reagents.

Atomics International/Wheelabrator-Frye

Under agreements with Atomics International, several other reagents were used with varying degrees of success. Reagents tested were potash, fly ash from several sources, limestone, slaked lime, hydrated lime, dolomitic lime, and ammonia. Initial test results indicated that slaked lime had promise as the most economical reagent.

In addition to the pilot testing done by Atomics International/Wheelabrator Frye -- other companies, primarily Joy Manufacturing Co./Niro Atomizer, Carborundum and Babcock & Wilcox pursued dry scrubber piloting.

Joy/Niro

The Joy Manufacturing Company and Niro Atomizer invested several million dollars to pilot a system in order to obtain data, optimize the process, and determine the economic feasibility of the dry scrubbing system. Niro had good experience in their Copenhagen test labs and in their pilot plants in Europe using both lime and sodium as reagents. From this, Niro made an appraisal and decision to proceed with a large scale pilot plant.

The Joy/Niro pilot was located at the Hoot Lake Station Unit 2 owned by Otter Tail Power Company at Fergus Falls, MN. Joy/Niro developed a recirculating system in which lime slurry is mixed with recirculated fly ash and spent reagent, for reinjection into the reactor, (patent pending).

Babcock & Wilcox

Babcock & Wilcox installed a semi-wet reactor designed on the basis of Japan's Hitachi process, followed by parallel streams to a precipitator and a baghouse at our William J. Neal Station at Velva, ND. Babcock & Wilcox later redesigned the system to a horizontal reactor using a "Y" jet dual fluid atomizer followed directly by an electrostatic precipitator. The flue gas and the slurry spray entered the reactor at the front wall, the geometry being very similar to that of a circular burner.

Carborundum

The Carborundum Company, using a De Laval spray dryer in combination with their baghouse, had a pilot test program on Unit 1 of our Leland Olds Station. They tested both dual fluid spray nozzles and rotary atomizers.

Test Parameters

All four pilot plants conducted tests which were based on the design conditions for our Antelope Valley Station. These parameters ranged from 400-2000 ppm SO₂ at the inlet, at approximately 310°F inlet flue gas temperatures, while burning North Dakota lignite.

Wyoming (Powder River Basin) sub-bituminous coal was later burned in the boilers used for the Joy/Niro and Babcock & Wilcox pilots. These test conditions were typical of the flue gas characteristics which will be encountered at the Laramie River Station, i.e. 400-800 ppm SO₂.

Reagents tested at the four pilot plants included soda ash, fly ash, lime, limestone, magnesium oxide and ammonia.

All pilots conducted one hundred hour endurance tests which demonstrated that the emission requirements were fulfilled with a comfortable safety margin.

With each week of testing, more and more was learned about the process. This included such parameters as lime slaking variables, feed slurry conditions, slurry feed temperatures, spray down temperatures, outlet gas temperatures, outlet humidity, atomizer design, atomizer speeds, spent reagent/ash recirculation and baghouse/precipitator SO₂ removal. Stoichiometric ratios improved to yield nearly 100% utilization of the lime feed.

DRY VERSUS WET SCRUBBERS

Although the process of dry scrubbing is new, the equipment is not. Spray drying technology is used extensively in mining applications and the food industry. Both fabric filter baghouses and electrostatic precipitators have been used for quite some time.

Some of the reasons for Basin Electric's enthusiasm for the dry scrubbers are as follows:

Waste Handling. The dry system has no sludge handling equipment, which is usually troublesome and has a record of high maintenance. Wet scrubbers require thickeners, centrifuges or vacuum filters and sludge-flyash blenders in order to obtain a dry product. The product from a dry scrubber can be handled with conventional dry handling systems used for flyash. For the coals tested, the dry product appears to handle as well as the flyash.

Wet/Dry Interface. Scaling and plugging is common in wet scrubbers at wet/dry interfaces and on scrubber packing materials and demisters. In the dry system, the interface point occurs in suspension; only dry powder makes contact with the walls. There are no packed beds or demisters in a dry scrubber.

Materials of Construction. Wet scrubbers require expensive alloy materials or coatings for protection from corrosion and erosion. The dry system can use low carbon steel for vessels and ductwork. The I.D. fans can be safely located just ahead of the stack without fear of fan corrosion and imbalance.

Operations. It is estimated that considerably fewer operations personnel will be required for the dry system. Wet scrubbers have proven to take considerable manpower for operations and maintenance. The dry scrubber offers flexibility of operation. Feed rates can be immediately adjusted with little concern for pH control. Turndown capability for a dry scrubber is in the order of a 10:1 ratio. Wet scrubber modules usually must be left in service at low loads to recirculate slurry. In the dry system, modules and/or atomizers can be removed from service quickly and easily as the load varies.

Maintenance. Wet systems have inherent high maintenance costs with slurry handling equipment recirculating abrasive materials at high pressures and volumes. The dry system operates with low pressures and low material volumes. Liquid to gas ratios are about 0.2 to 0.3 gallons per 1000 actual cubic feet compared to about 40-100 for a wet scrubber. The atomizer, which is probably the highest wear item in the system, can be removed and replaced quickly. The elimination of dewatering equipment reduces the maintenance expense considerably from that for a wet scrubber.

Energy Requirements. The dry system requires approximately 25% to 50% of the energy required for a wet system.

Particulate Collection. The gas volume to the particulate collector is reduced to below that leaving the air heater as a result of the spray down in the dry scrubber. The gas temperature to the particulate collection device remains constant at all loads. The spray dryer conditions the flyash with the added moisture resulting in a lower resistivity ash in the precipitator. Although the scrubber reactants produce additional particulates, the increased humidity and lower temperatures have a positive effect on precipitator operation. The pilot test results indicated that the increased humidity did not adversely affect baghouse operations, and the baghouse benefits from the reduced gas volume.

Water Consumption. The water requirements for a dry system are much less than for a wet system. The dry scrubber at the Laramie River Station Unit 3 will use about 50% of the amount required for the wet scrubber on Unit 1 or Unit 2. Low quality water such as cooling tower blowdown or ash water may be used in the spray dryer. Only a small quantity of treated water, about 20% of the total requirement, is needed for lime slaking.

ANTELOPE VALLEY STATION DRY SCRUBBER

After analyzing and evaluating the data from the four pilot plants with the help of our architectural engineer, Stearns-Roger, we took bids for a dry system utilizing lime for our Antelope Valley Station Unit 1. After receiving bids from four vendors, a contract for the Antelope Valley Station Unit 1 SO₂ and particulate removal system was awarded to Joy Manufacturing Company, Western Precipitation Division with Niro Atomizer Company as prime subcontractor. This plant is scheduled to begin commercial operation in April, 1982.

The system will have five vertical Niro spray dryers (four in operation, one as a spare) followed by a Western Precipitation baghouse (See Figure 1). Each spray dryer will utilize one rotary atomizer with a direct drive motor (See Figure 2). Quick lime will be slaked in a ball mill slaker. Lime slurry, sludge from the primary water treatment process, and a portion of the recycled ash product will be mixed to create the feed slurry which goes to the atomizers (See Figure 3). A dispersed mist of fine droplets is sprayed into the flue gas as it enters the spray dryer at the top and center. A very rapid chemical reaction occurs removing a substantial amount of SO₂ from the flue gas. At the same time, the thermal energy of the flue gas evaporates the water in the droplets to form a dry powder consisting of calcium sulfite, calcium sulfate, unreacted lime, and flyash.

Some of the product is removed in the conical bottom of the dryer. The remainder is suspended in the flue gas which goes to the baghouse (See Figure 4). Additional SO₂ is removed by the powder on the fabric filter bags, thus increasing the system efficiency.

The baghouse has 28 compartments having a total of approximately 8,000 fiberglass filter bags with fluoro-carbon coating. The bags are 12 inches in diameter and 35 feet high. Reverse air cleaning will be used. The gas to cloth ratio will be 2.19 cubic feet of gas per minute per square foot of cloth (gross) under maximum operating conditions.

The cleaned flue gas exits the baghouse to the I.D. fans and is exhausted to the stack.

The SO₂ and Particulate Removal System is designed to operate at 62% SO₂ removal for performance lignite (average sulfur) and 78% SO₂ removal for design lignite (maximum sulfur) in order to meet the emission limitation of 0.78 lb. SO₂ per million Btu as required by the North Dakota State Department of Health. The coal analysis, is given in Table 1, and the operating conditions and flue gas characteristics are given in Table 2.

The lime feed rate is essentially stoichiometric. A low stoichiometric ratio is possible because of the utilization of available alkalinity in the flyash.

Antelope Valley Station Gas Cleaning System

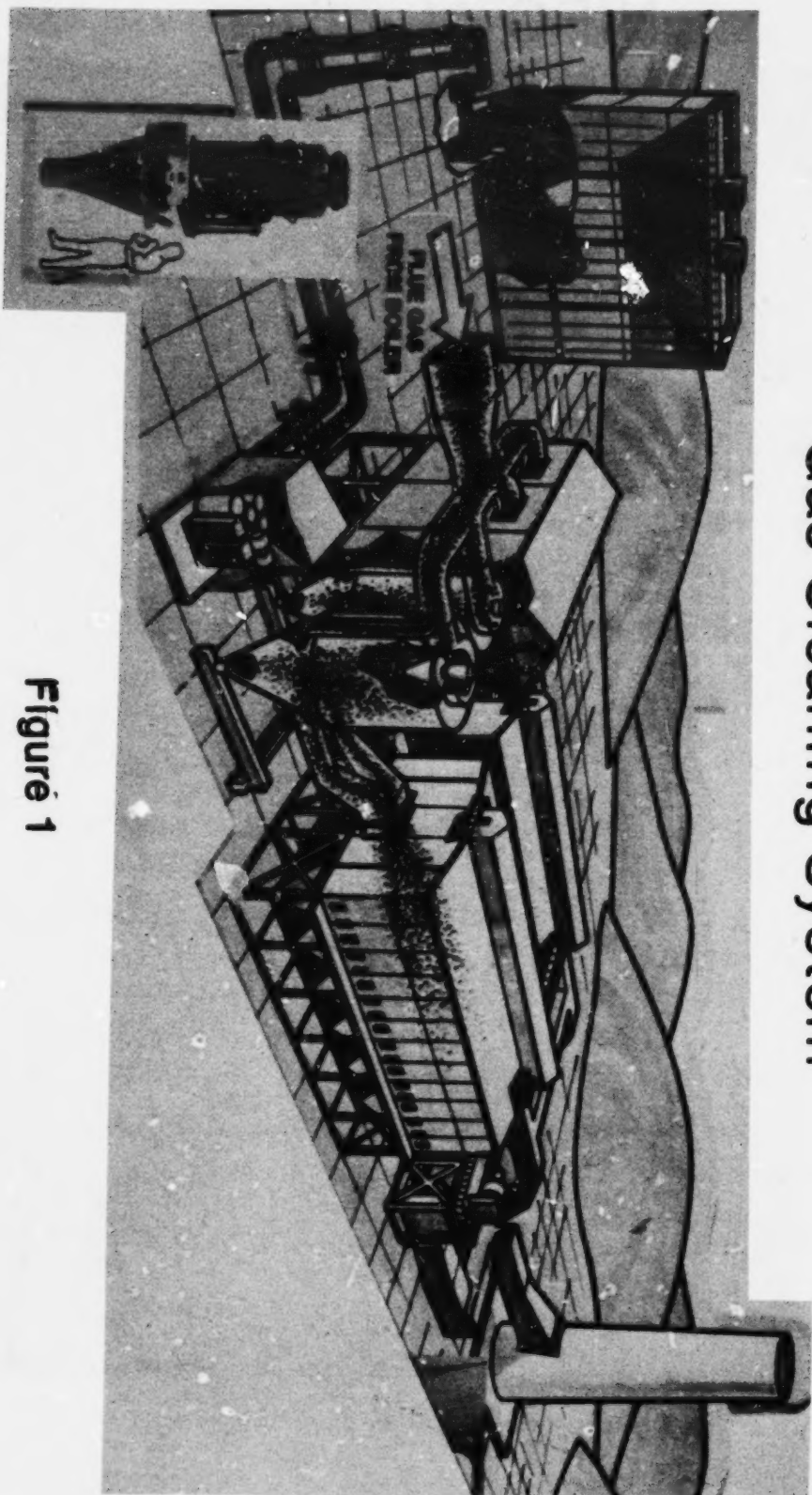


Figure 1

NIRO Atomizer

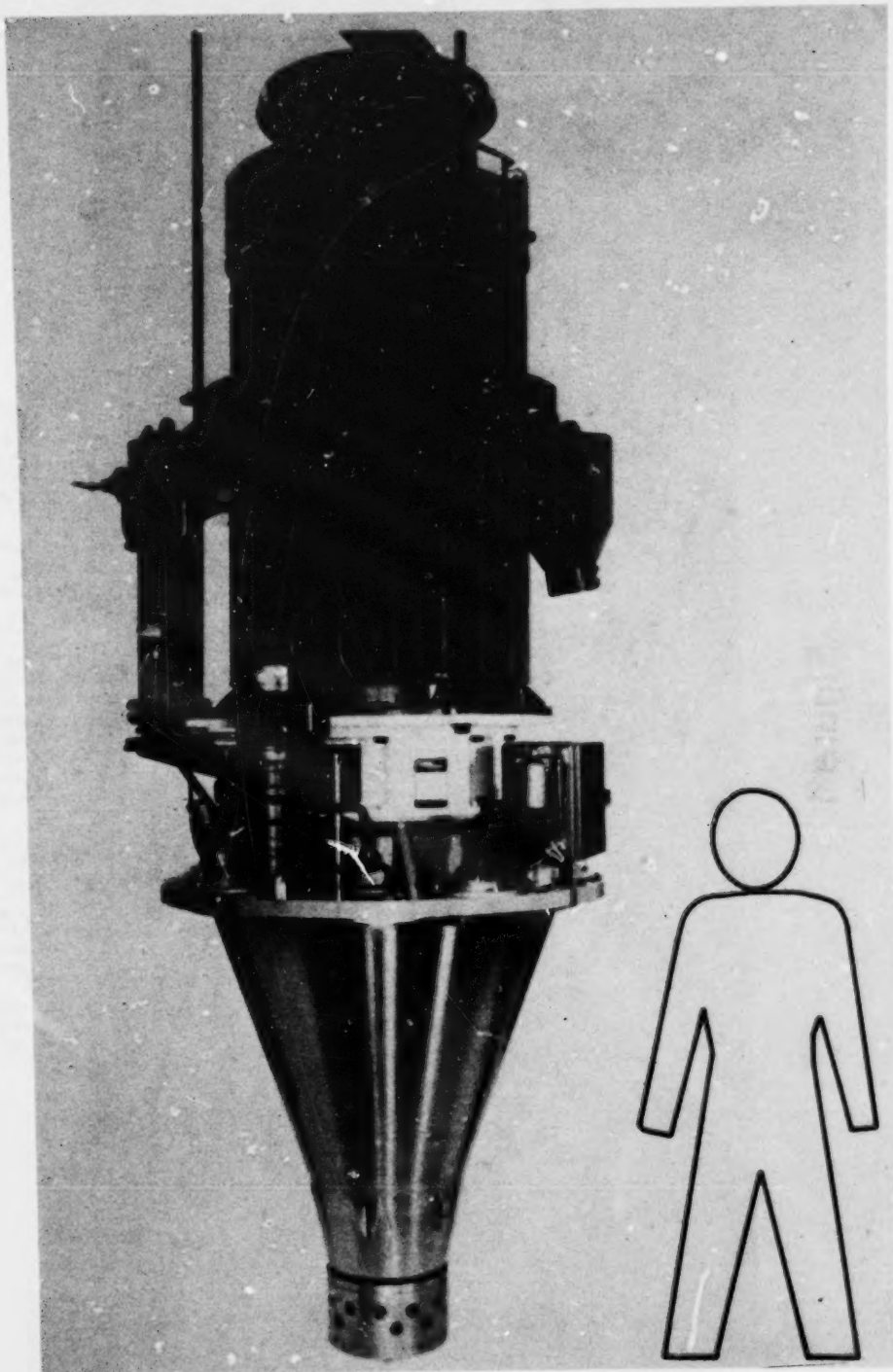


Figure 2

Antelope Valley Station Flow Diagram

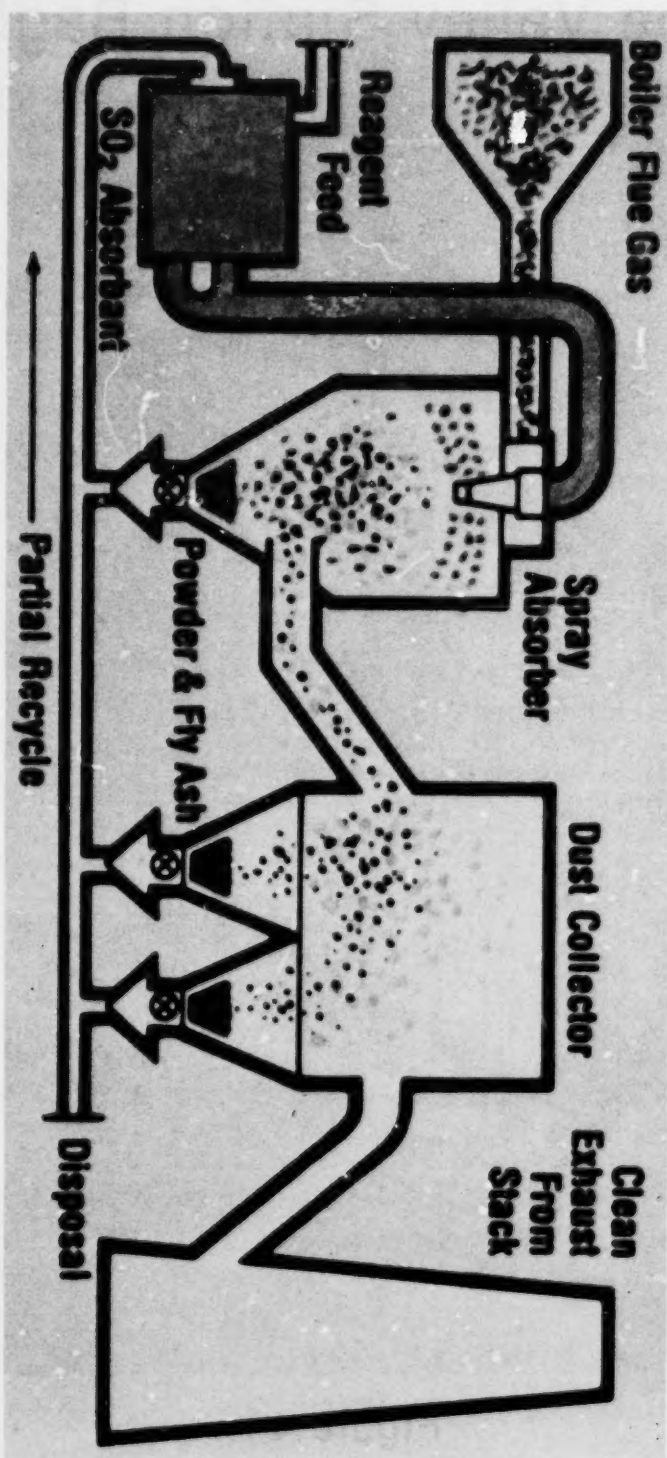


Figure 3

Antelope Valley Station Baghouse

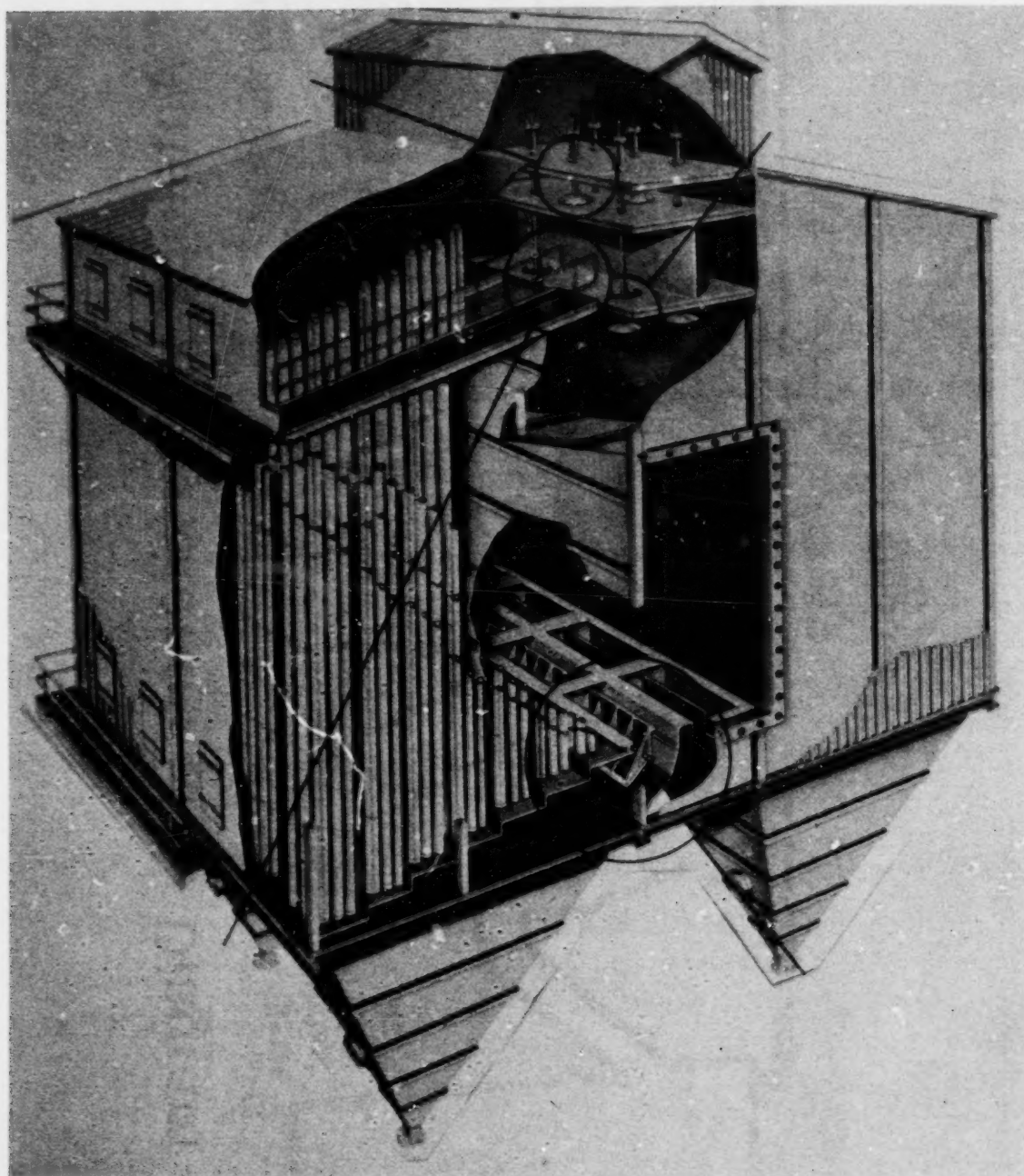


Figure 4

Table 1. ANTELOPE VALLEY STATION
FUEL CHARACTERISTICS
ULTIMATE ANALYSIS

	Average %	Range %
Carbon	39.69	36.55 - 41.92
Hydrogen	2.70	2.38 - 2.96
Nitrogen	0.60	0.45 - 0.81
Sulfur	0.68	0.36 - 1.22
Moisture	37.07	30.03 - 42.43
Oxygen	11.49	10.15 - 12.86
Chlorine	0.01	0.00 - 0.02
Ash	7.76	5.24 - 13.83
HHV, Btu/lb	6600	6093 - 7350

Table 2. ANTELOPE VALLEY STATION
OPERATING CONDITIONS
&
FLUE GAS CHARACTERISTICS

OPERATING CONDITIONS		
Generation Net/Gross, MW	385/440	
Fuel Input	4930x10 ⁶ Btu/hr	
Coal Burn Rate	375 tons/hr	
FLUE GAS CHARACTERISTICS		
	INLET	OUTLET
Gas Flow, acfm	2,055,000	1,894,380
scfm	1,248,900	1,369,000
Gas Temperature °F	310	185
Mass Gas Flow, lbs/hr	5,690,000	6,097,000
Moisture, % Volume	15.6	19.4
SO ₂ ppm by Volume	800	304
lbs/hr	10,120	3,845
lbs/MKB	2.07	0.78
Particulate lbs/hr	88,070	210
gr/acf	5.0	0.012
gr/scf	7.7	0.018

Based on anticipated average fuel analysis and 100% plant load.

LARAMIE RIVER STATION DRY SCRUBBER

We had Burns & McDonnell prepare specifications and take bids for a dry system for the Laramie River Station Unit 3. These bids were evaluated and the economics compared to that of an option available for a precipitator and a wet scrubber identical to Units 1 & 2. A contract for the Gas Cleaning System for Unit 3 of the Laramie River Station was awarded to Babcock & Wilcox. Unit 3 is scheduled to begin commercial operation in April, 1982. This scrubber will be a variation of the concept used at the Antelope Valley Station. The system consists of four reactors (three in operation, one as a spare) each followed by an electrostatic precipitator (See Figure 5).

Each Babcock and Wilcox reactor is equipped with 12 "Y" jet nozzles. The dual fluid atomizers will use a concentric pipe feeder for the atomizing fluid, which is steam, and the reagent, which is lime slurry. (See Figure 6). The atomizer is horizontally fitted into the center of a circular throat on the front wall of the reactor (See Figure 7). The throat is fitted with vanes which control the shape of the gas envelope. The angle of the atomizer spray cone is matched to this envelope to obtain the maximum mixing of the atomizer slurry and the flue gas. This principle has been used in the mixing of fuel and air since the 1930's.

Under each reaction chamber are three hoppers to collect a portion of the product which will drop out prior to the precipitator (See Figure 8).

A portion of flue gas (approximately 3%) bypasses the air heater and reactor and enters the reactor discharge plenum where it is mixed with the reactor discharge for reheat. The reheated flue gas enters the B & W Rothemule electrostatic precipitator, which is the same design and size as that on Unit 1 and Unit 2.

The gas cleaning system is designed to operate at 85% SO₂ removal with a 0.54% sulfur coal. The system is also designed for 90% SO₂ removal with a 0.81% sulfur coal. The coal analysis is given in Table 3, and the operating conditions and flue gas characteristics are given in Table 4.

WASTE PRODUCT DISPOSAL

The waste product material has an appearance of being totally dry. There is a relatively small amount of moisture present, most of which is chemically bound water of hydration. The waste product is a very fine grain, powdery material, very similar in particulate size distribution as the flyash normally removed from the flue gas stream of a coal-fired boiler. A summary of physical characteristics from a waste product sample, selected from a pilot plant burning North Dakota lignite and using lime as the reagent, is given in Table 5. Of particular significance is the relatively impermeable nature of the material.

The amount of water soluble material found in the spray dryer product is dependent on the reagent material--the relative solubility of

Basin Electric Power Cooperative **Laramie River Station - Unit No. 3**

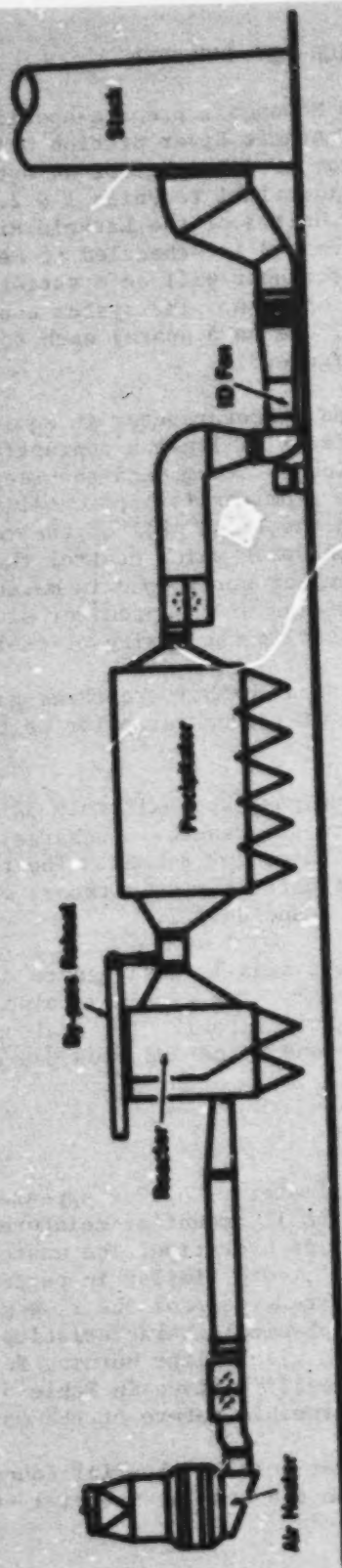
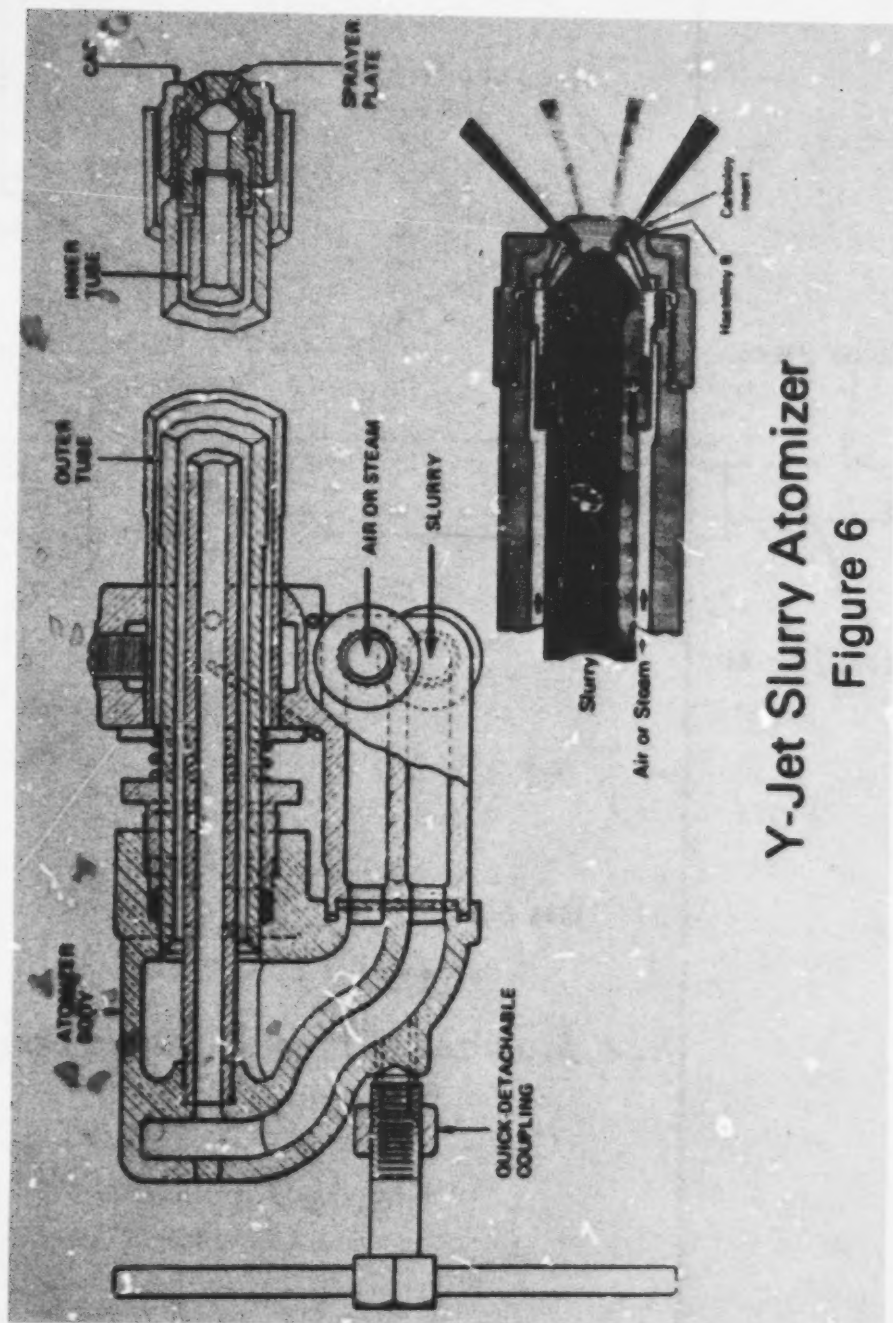
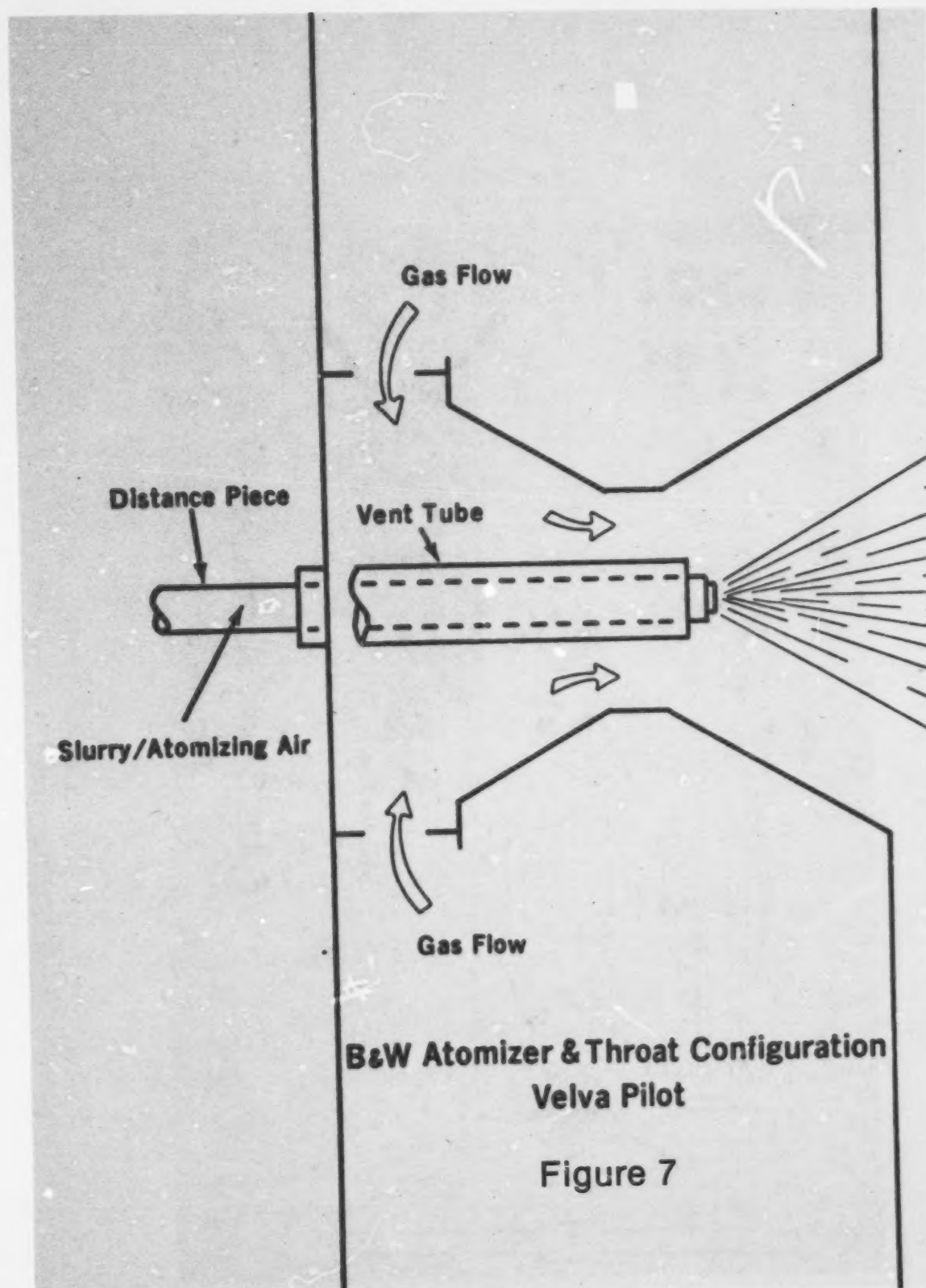


Figure 5



Y-Jet Slurry Atomizer
Figure 6



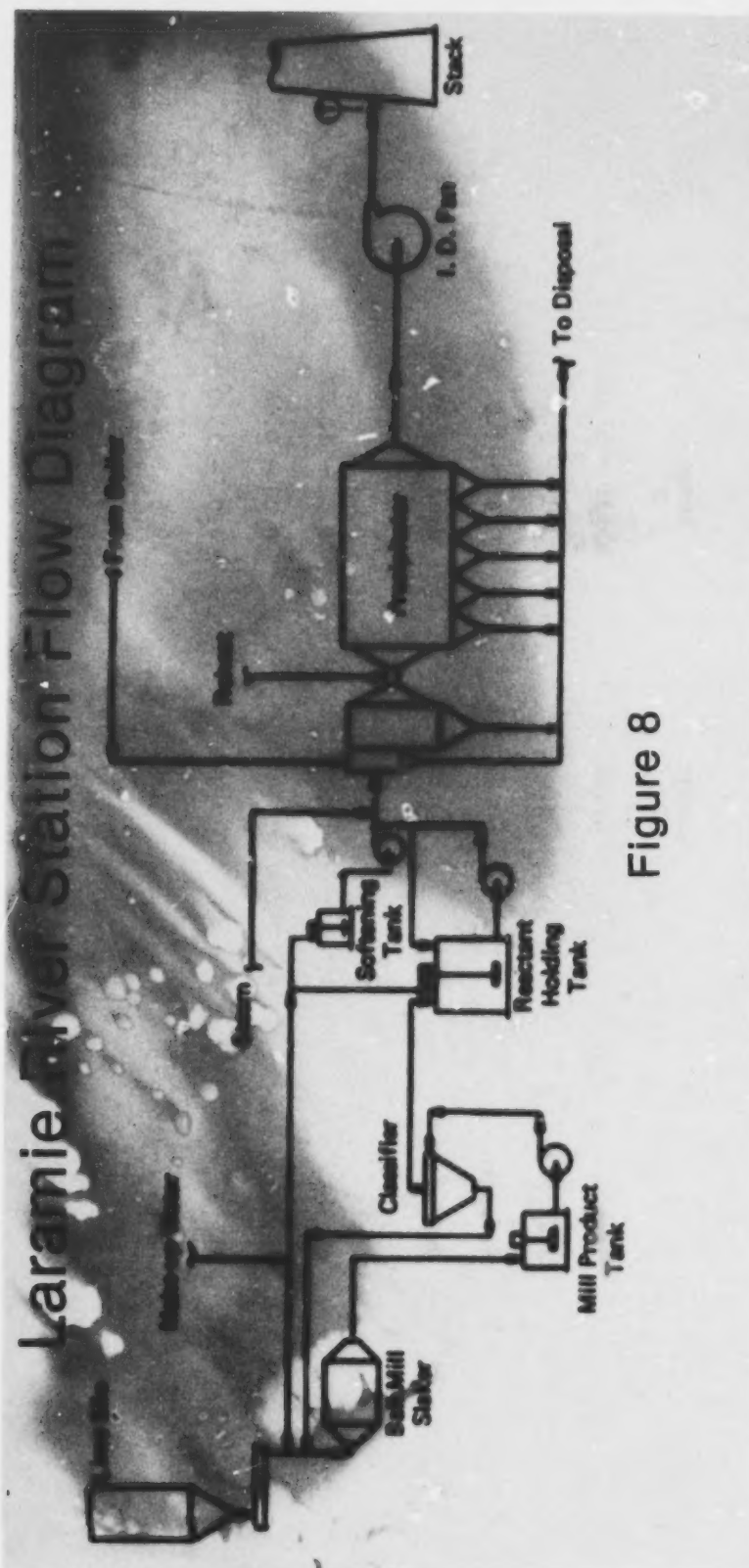


Figure 8

Table 3. LARAMIE RIVER STATION
FUEL CHARACTERISTICS
ULTIMATE ANALYSIS

	Average %	Range %
Carbon	49.07	45.32 - 51.32
Hydrogen	3.45	3.03 - 3.91
Nitrogen	0.92	0.4 - 1.0
Sulfur	0.54	0.15 - 0.54
Moisture	28.92	22.2 - 34.65
Oxygen	9.17	8.51 - 12.07
Chlorine	0.04	0.00 - 0.04
Ash	7.89	4.0 - 13.0
HHV, Btu/lb	8139	7906 - 8244

Table 4. LARAMIE RIVER STATION
OPERATING CONDITIONS
&
FLUE GAS CHARACTERISTICS

OPERATING CONDITIONS		
Generation Net/Gross, MW	500/575	
Fuel Input	5513.625x10	Btu/hr
Coal Burn Rate	340 tons/hr	
FLUE GAS CHARACTERISTICS		
	INLET	OUTLET
Gas flow, acfm	2,300,000	1,946,000
scfm	1,320,000	1,405,000
Gas Temperature °F	286	157
Mass Gas Flow, lbs/hr	6,188,000	6,461,000
Moisture % Volume	11	17
SO ₂ ppm by volume	530	80
lbs/hr	7320	1100
lbs/MKB	1.33	0.2
Particulate lbs/hr	78,860	300
gr/acf	4.0	0.015
gr/scf	7.0	0.02

Based on anticipated average fuel analysis and 100% plant load.

Table 5. PHYSICAL CHARACTERISTICS OF WASTE PRODUCT

Specific Gravity	2.8 - 2.9
Maximum Compacted Dry Density	62 pcf
Optimum Moisture Content	55%
Permeability	1×10^{-6} to 1×10^{-7} cm/sec
Atterberg Limits	
Liquid Limit (%)	63
Plastic Index (%)	12

calcium sulfite and calcium sulfate as compared to sodium sulfite and sodium sulfate. In this respect the solubility of a sodium based product has been found to be 50-60% soluble, whereas the solubility of a calcium based product has been measured at 3-7% solubility. The ratio of sulfite to sulfate in the waste product is quite variable. In sodium based systems, sulfate is by far the most predominate specie. In lime based systems, the ratio is more balanced. (1:1 to 2:1 favoring sulfite). In addition to sulfite and sulfate as spray dryer reactants, there is also some unutilized reagent present in the product, and also there is a small amount of carbon dioxide absorption from the flue gas which forms calcium carbonate in the product. Table 6 gives the expected chemical characteristic of the waste product from the Antelope Valley Station.

The method of waste disposal planned for the Antelope Valley Station is to transport the waste product to depleted areas of the mine for landfill. The method of waste disposal planned for Laramie River Station is to transport the material to a landfill where it will be disposed of along with conventional limestone scrubber sludge.

The material is fairly cementitious and impermeable, however, potential problems in the field might be weathering, erosion, dust suppression and structural stability. Disposal procedures will be determined during field tests when the product first becomes available.

It is germane to note that there are no apparent disadvantages in disposing of a calcium based scrubber waste product compared to that of flyash. A waste product from a sodium based system may require special handling, however, because of the increased solubility of the sodium sulfite and sodium sulfate.

We believe there may be a potential market for the sale of the waste product.

ECONOMICS

The economics of any FGD or particulate control technology for full-scale utility boilers are unquestionably site-specific. Economics are subject to wide variations depending on the type of fuel burned, applicable environmental standards and geographical location of the power plant. However, as a result of extensive engineering studies, the operational experience with full-scale spray dryers, baghouses and precipitators, and the many pilots tests performed to date; the general economics can be derived with a relatively high degree of confidence. The economic analysis of wet versus dry systems for both the Antelope Valley Station and the Laramie River Station gave the dry system the lowest evaluated cost over the lifetime of the plants.

Evaluated costs for the dry scrubber system at the Antelope Valley Station produced approximately \$47 million cost savings over a wet scrubber system for a plant life of 35 years based on present worth and a 75% annual plant factor (See Table 7). This is attributed to a savings in operation and maintenance expenses as evaluated by Basin Electric.

Table 6. EXPECTED WASTE PRODUCT CHEMICAL CHARACTERISTICS
FROM THE ANTELOPE VALLEY STATION

SO ₂	Reactants	%	Fly Ash	%
	CaSO ₄	8.9	SiO ₂	21.1
	CaSO ₃	14.6	P ₂ O ₃	0.4
	CaCO ₃	2.1	Fe ₂ O ₃	6.8
	Lime Inerts	1.6	Al ₂ O ₃	8.5
	H ₂ O	1.3	TiO ₂	0.5
	Total	28.5	CaO	18.1
			MgO	5.8
			Na ₂ O	4.7
			K ₂ O	0.5
			SO ₃	3.4
			Undetermined	1.7
			Total	71.5

Table 7. ANTELOPE VALLEY STATION
WET VS. DRY SCRUBBER
ECONOMIC EVALUATION

ECONOMIC FACTORS	DRY SCRUBBER		WET SCRUBBER	
	35 Year Total Cost \$1981	Basis of Evaluation	35 Year Total Cost \$1981	Basis of Evaluation
Capital Cost	\$49,665,100	actual	\$55,927,400	estimate
Reagent Cost	\$38,587,500	18,000 tons/yr lime	\$42,505,000	41,848 tons/yr limestone
Power Cost	\$ 7,466,700	5726 KW	\$13,040,000	10,000 KW
Manpower Cost	\$15,925,000	6 operators (est.) 7 maintenance (est.)	\$36,750,000	11 operators (est.) 19 maintenance (est.)
Replacement Materials	\$17,500,000	filter bags, atomizer wheel, bearings, etc. (est.)	\$28,000,000	pH control, sludge fixation chemicals, pump misc., ESP bushings, & TR's (est.)
Total	\$129,144,300		\$176,222,400	

Evaluation based on 35 year life, annual plant factor of 75% and present worth of operation and maintenance cost as of the unit's commercial date (1982).

There are intangibles with both the wet system and the dry system that are difficult to place a dollar value on. If a wet system had been selected for the Laramie River Station Unit 3, such intangibles as reduced spare parts and maintenance know-how of similar systems would have been realized. Yet, in the dry system, there are potential improvements that may be obtained in stoichiometry as refinements are made to the atomization, mixing techniques, and the slaking process. A substantial improvement can be made in the economics if lime becomes locally available. Limestone is locally available, but the closest source of lime is approximately 200 miles.

Even though these intangibles are not evaluated, the total evaluated cost of the wet system is still higher than the dry system. This is true even after considering the cost of transporting lime to the site versus using locally available limestone. At the Laramie River Station Unit 3, the evaluated costs of the dry system are approximately \$23 million less than the wet system based on a plant life of 35 years and a 75% annual plant factor (See Table 8).

CONCLUSION

In the long run, availability will probably be the most important advantage of the dry system. Attempts will not be made here to develop numbers for this, since it will be three to four years before operating data will be available. However, it should be apparent that the dry system, when compared to wet scrubbers, has less equipment which can cause trouble. The dry system has fewer pieces of complex equipment, is simpler to control, and is not particularly sensitive to changes in operating conditions.

We believe the utility industry will eventually accept the concept of dry scrubbing, not so much due to the economic factor, but because of the system's simplicity and potential availability. It is only fair to say that the figures quoted comparing the wet and dry systems are site specific for our applications. Changes which would tend to favor the economics of the wet system would be an increase in the differential between the cost of lime and limestone and an increase in the coal sulfur content. Although it is beyond the scope of this presentation, results from continuing developmental work are increasingly favorable, appearing to make the dry system competitive on higher sulfur coals. Changes which would favor the dry system would be: (a) improvement in system stoichiometry, (b) improvement in availability of the dry scrubber compared to the wet scrubber, (c) future escalation in material and labor costs, (d) higher fixed charges and (e) increased energy costs. All of these appear to be most likely probabilities.

While the dry scrubber is new, we believe that it will be the scrubber of the future, particularly for low sulfur western coals.

Table 8. LARAMIE RIVER STATION
WET VS. DRY SCRUBBER
ECONOMIC EVALUATION

ECONOMIC FACTORS	DRY SCRUBBER		WET SCRUBBER	
	35 Year Total Cost \$1981	Basis of Evaluation	35 Year Total Cost \$1981	Basis of Evaluation
Capital Cost	\$49,807,000	actual cost + ancillaries	\$60,632,100	estimate
Reagent Cost	\$48,880,000	20,920 tons/yr lime	\$ 9,367,000	35,400 tons/yr limestone
Power Cost	\$ 4,806,000	2451 KW	\$19,610,000	10,000 KW
Manpower Cost	\$15,925,000	6 operators (est.) 7 maintenance (est.)	\$36,750,000	11 operators (est.) 19 maintenance (est.)
Replacement	\$15,750,000	spray nozzles, pump misc., ESP bushings & TR's (est.)	\$28,000,000	pH control, sludge fixation chemicals, pump misc., ESP bushings, & TR's (est.)
Pressure Loss	\$ 4,623,800	6.5" w.g., air heater outlet to chimney inlet	\$ 8,678,600	12.2" w.g., air heater outlet to chimney inlet
Total	<u>\$139,791,800</u>		<u>\$163,037,700</u>	

Evaluation based on 35 year life, annual plant factor of 75% and present worth of operation and maintenance cost as of the unit's commercial date (1982).

UTILITY CONVENTIONAL COMBUSTION COMPARATIVE
ENVIRONMENTAL ASSESSMENT - COAL AND OIL

Charles A. Leavitt, C. Shih, Rocco Orsini, Kenneth Arledge, and Alexandra Saur
TRW, Inc., Redondo Beach, Cal.

Warren D. Peters
U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, N.C.

ABSTRACT

This paper summarizes the results of a comparative multimedia assessment of two utility boilers, one firing coal, the other oil, to determine relative environmental impacts. Comprehensive sampling and analyses of multimedia emissions from the boilers and control equipment were conducted to identify criteria pollutants and other species. The results indicate that: (1) particle emissions from coal firing are 10 times those from oil firing (scrubbing reduced the mean size of the particles from coal firing and increased the emissions of particles below 3 μm size); (2) SO_2 emissions from coal firing are 7 times those from oil firing; (3) NO_x emissions from coal firing are 6 times those from oil firing; (4) for coal firing, arsenic, cadmium, chromium, iron, nickel, lead, and zinc emissions are of potential environmental concern -- for oil firing, only chromium and nickel are of potential concern; (5) organic and polycyclic organic emissions from either fuel are not of environmental concern; (6) liquid emissions from coal firing contain many inorganic elements that are of environmental concern although the organic content is innocuous; and (7) solid emissions from coal firing also contain many trace elements at concentrations of concern -- there are no significant solid wastes from oil firing.

INTRODUCTION

Objective

Conventional methods of converting fossil fuels to usable forms of energy have impacts on the air, land, and water; i.e., multimedia impacts. These impacts are not separate and distinct; rather, they are all inter-related and involve delicate balances and trade-offs.

The Environmental Protection Agency (EPA), with primary responsibility for controlling adverse environmental impacts of pollutant emissions, has been active since its inception in determining the identities and quantities of potential pollutants released to the environment when fossil fuels are burned. Information from EPA R&D efforts is being used for three principal purposes: to assess the health and environmental impacts caused by the release of combustion pollutants to the environment; to define the needs for technology to control the release of these pollutants; and to develop standards to limit emissions.

CCEA Program

In response to the need for a comprehensive environmental assessment of conventional combustion systems, EPA's Industrial Environmental Research Laboratory at Research Triangle Park (EPA/IERL - RTP), North Carolina, has established a unified Conventional Combustion Environmental Assessment (CCEA) program.¹ It is a major new program aimed at the comprehensive assessment of environmental, economic, and energy impacts of multimedia pollutant emissions from stationary industrial, utility, residential, and commercial conventional combustion processes. The primary objective of the CCEA program is to identify and evaluate information from all relevant sources in order to: determine the extent to which this information can be utilized to assess the total environmental, economic, and energy impacts of conventional combustion processes; identify and acquire additional information needed for such assessment; define the requirements for modifications or additional development of control technology; and define the requirements for modified or new standards to regulate pollutant emissions.

The CCEA program will coordinate and integrate ongoing and future studies into a unified environmental assessment structure and serve as a centralized base of information on the environmental impacts of conventional combustion processes. Coordination and information exchange between CCEA-related studies should minimize duplication and maximize the return from available resources.

The environmental assessment (EA) methodology employed in the CCEA program draws heavily on the philosophies of the existing EPA/IERL-RTP EA methodology, but has been expanded and modified to be more responsive to the assessment of conventional combustion processes.

In the most elementary description, an environmental assessment consists of three basic iterative steps (Figure 1):

1. Characterization of the combustion process (including any associated pollution control devices) and its effluents.
2. Assessment of the health and ecological impacts of the combustion process and its effluents on the environment.
3. Evaluation of alternative control strategies to reduce pollution impacts to acceptable levels.

The EA procedure used in the CCEA program is shown in the generalized methodology diagram (Figure 2).

It is the goal of the CCEA program to integrate ongoing projects and recommend new efforts to address all practical combinations of information. It is expected that EPA/IERL-RTP, with the assistance of contractors with experience and expertise in the various areas associated with the comprehensive environmental assessment of conventional combustion processes, will implement and expand the CCEA program as needs dictate and as resources permit.

Multimedia Assessment Method

A major goal of the CCEA program is to support the implementation of the National Energy Plan, which aims at increasing the use of coal. Since fuel switching from oil to coal is an important facet of the NEP, the CCEA program initiated a study to compare the environmental impacts of oil and coal combustion.

The objectives of this study were to conduct multimedia environmental assessments of oil and coal firing in a controlled utility boiler in order to compare environmental, energy, and societal impacts of firing coal vs. firing oil. In order to conduct the comparative assessment, it was necessary to fully characterize feed streams, emissions, and effluents from the utility boilers selected for study and all associated pollution control equipment.

Plant Description and Fuel Analyses

Two electrical utilities were chosen for this study, with their agreement and cooperation. A midwestern coal fired utility with flue gas desulfurization was one selection. A west coast oil fired utility with staged combustion and flue gas recirculation was the other selection.

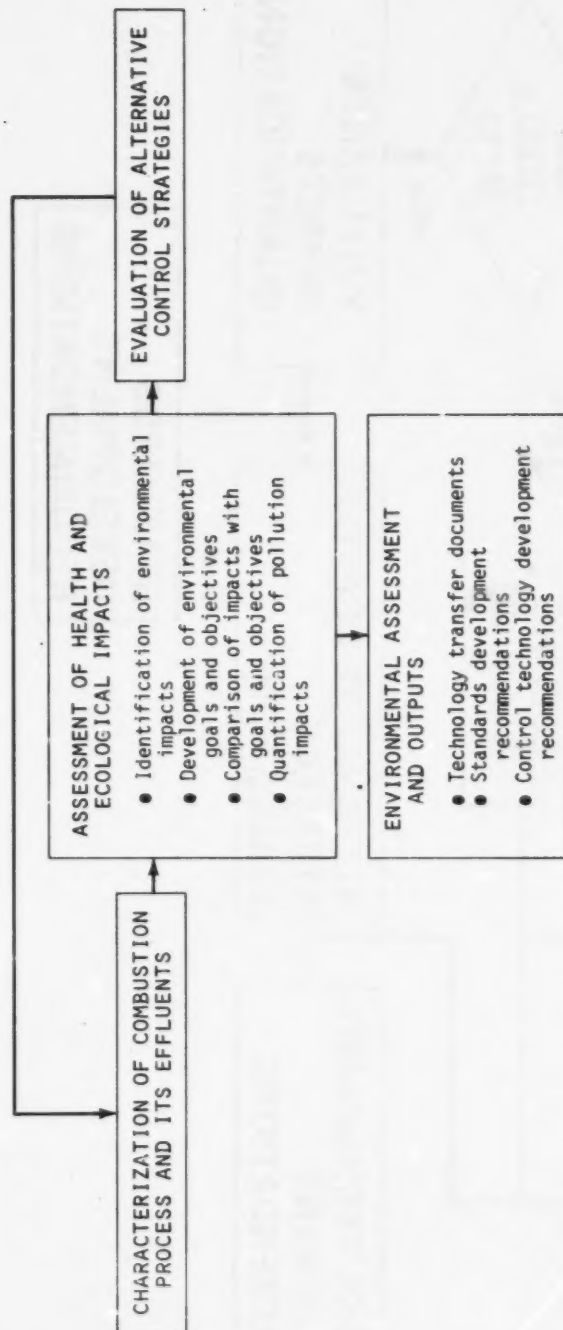


FIGURE 1
GENERALIZED ENVIRONMENTAL ASSESSMENT METHODOLOGY

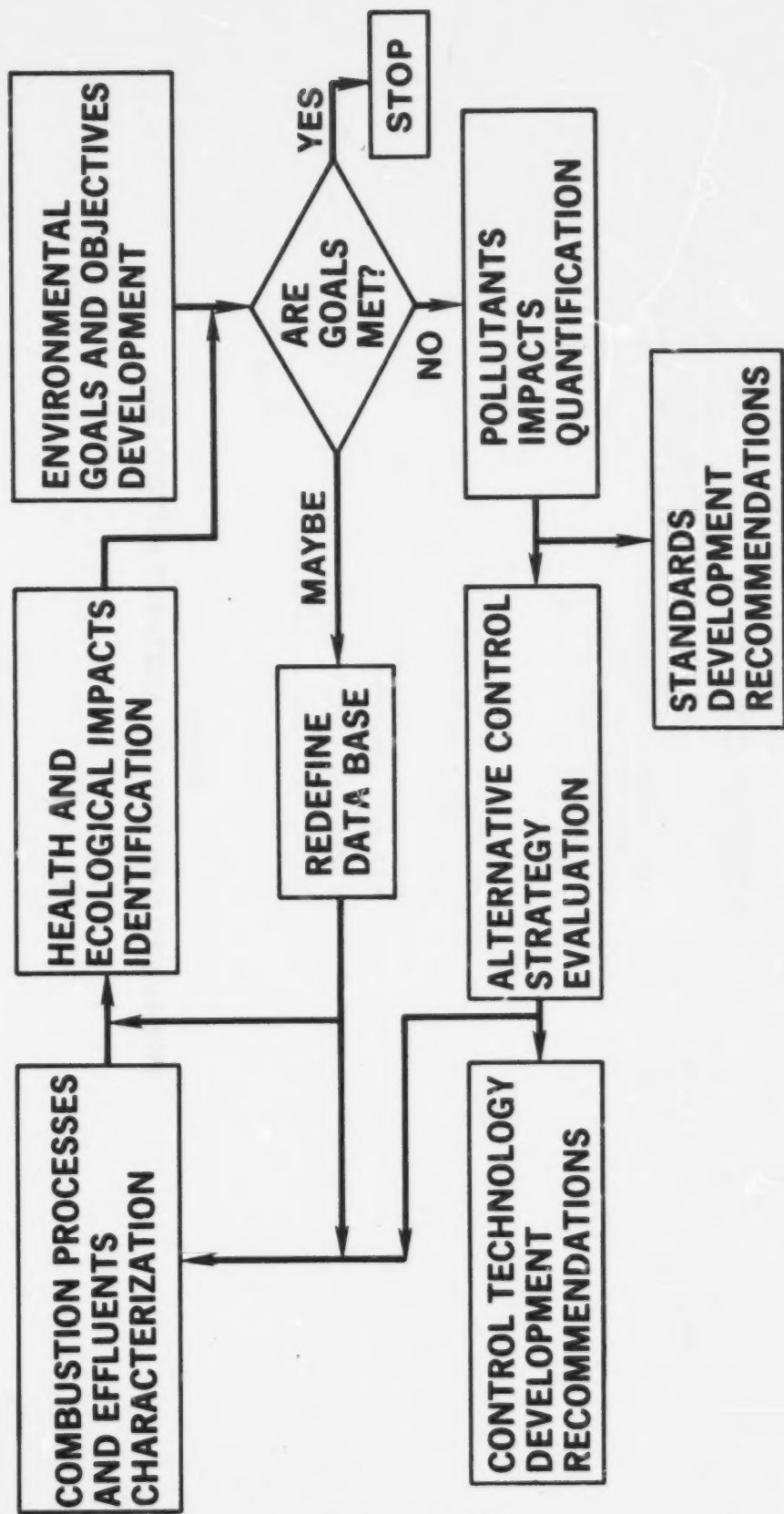


FIGURE 2
GENERAL METHODOLOGY

The coal fired utility boiler tested burns local coal and is equipped with an emission control system composed of eight two-stage venturi/absorber scrubber modules using a slurry of local limestone, see Figure 3. The boiler is a cyclone fired, supercritical, once through, balanced draft B&W unit rated at 2.81×10^6 kg steam per hour at 538°C, 26.13 MPa (6.2×10^6 lb/hr at 1000°F, 325 psig). The coal is a low grade sub-bituminous class with a typical as-fired heating value of 20.9 to 22.6 MJ/kg (9000 to 9700 Btu/lb), with an ash content of 25 percent and containing 5 to 6 percent sulfur, see Tables 1 and 2.

TABLE 1.

COAL FIRED BOILER DESIGN DATA

Type	Crushed coal, cyclone fired
Manufacturer	Babcock and Wilcox
Type of Burner	Cyclone
Number of Burners	18
Air Preheaters	Yes
Fuel	Sub-bituminous from Pittsburgh and Midway Coal Mining Company
Design Steam Rate	2.81×10^6 kg/hr (6.2×10^6 lb/hr) 26.13 MPa (325 psig), at 538°C (1000°F)

The oil fired utility boiler burns low sulfur fuel oil and utilizes sub-stoichiometric firing with flue gas recirculation to control NO_x emissions. No other air quality control equipment is installed, see Figure 4. Boiler No. 5 was tested. It was designed to burn either natural gas or oil. Oil analyses are given in Table 2. It is a double reheat, supercritical pressure, once through, front and rear fired B&W Universal Pressure boiler. See Table 3.

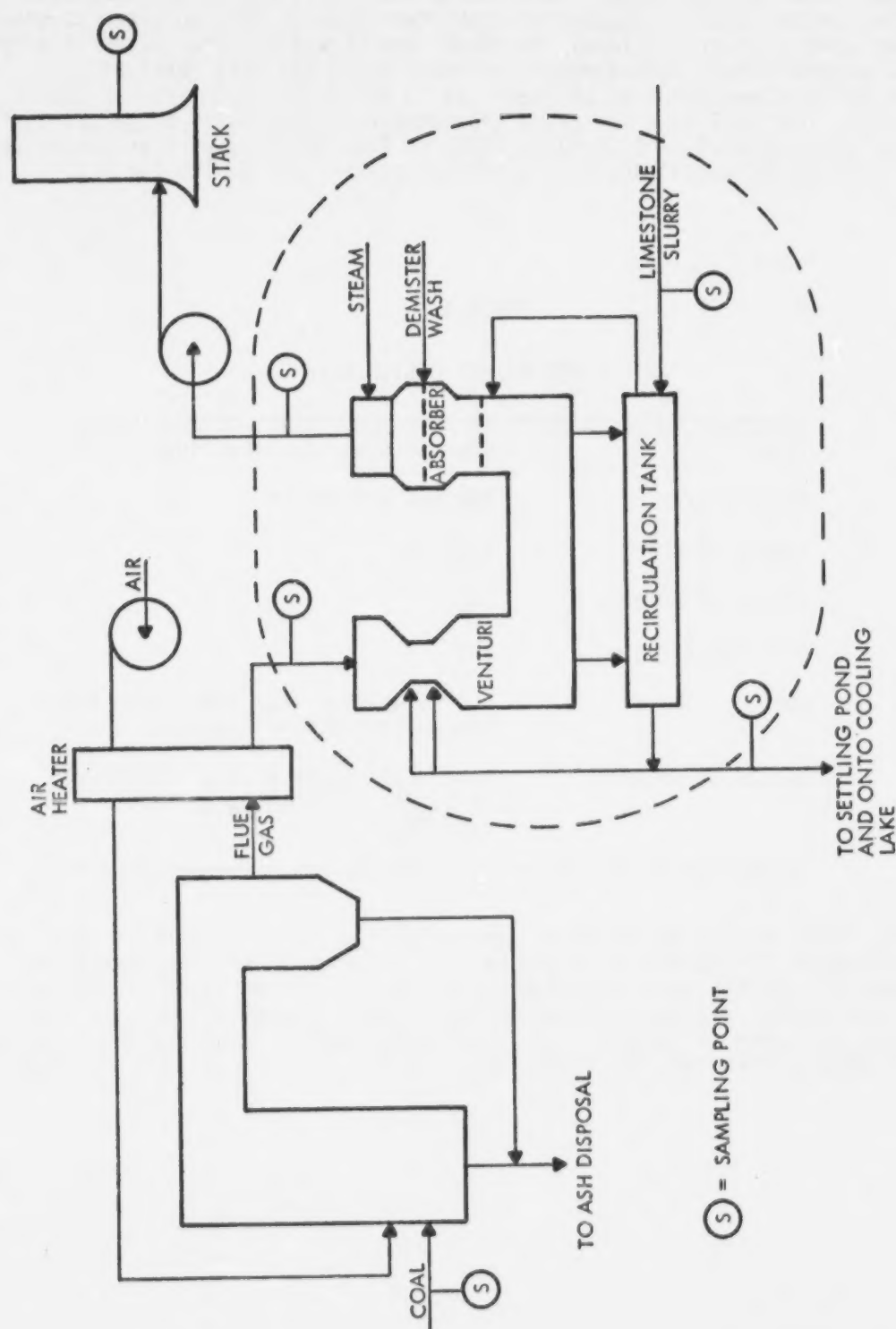


FIGURE 3
COAL FIRED BOILER SCHEMATIC

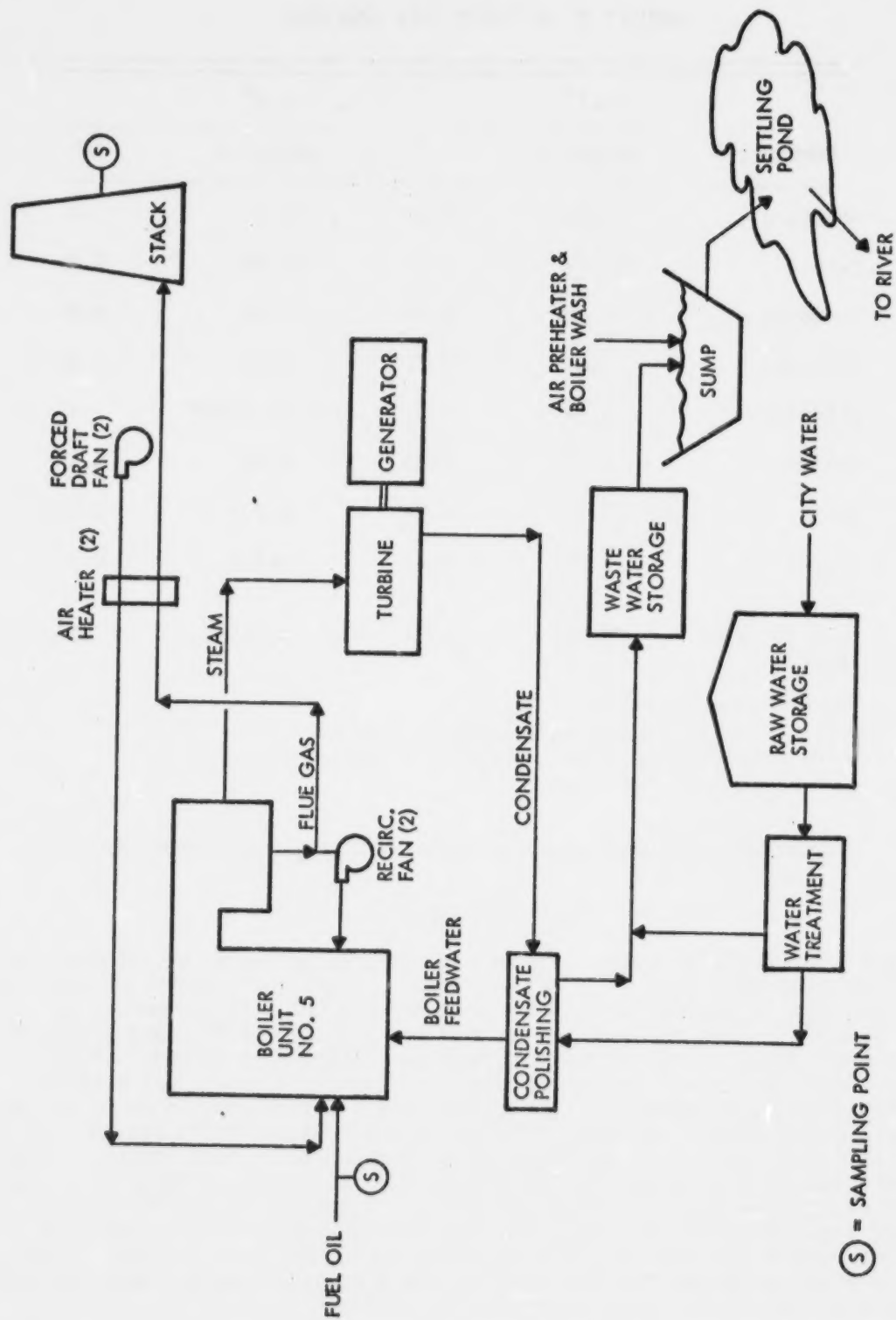


FIGURE 4
OIL FIRED BOILER SCHEMATIC

TABLE 2.
SUMMARY OF ULTIMATE FUEL ANALYSES

Component	Coal ^a		Oil ^b	
	Weight %	σ^c	Weight %	σ^c
Moisture	1.34	0.07	~ 0	--
Carbon	57.28	1.28	86.54	0.31
Hydrogen	3.84	0.09	12.39	0.04
Nitrogen	1.01	0.12	0.39	0.33
Chlorine	0.04	0.01	Not Analyzed	--
Sulfur	5.45	0.64	0.81	0.02
Ash	26.56	1.03	0.008	0.005
Oxygen	4.48	0.66	0.658	0.41
Heating Value (kj/kg)	24,027	384	44,025	103

a Coal analysis based on average of 5 tests.

b Oil analysis based on average of 4 tests.

c σ = standard deviation.

Control Devices

The coal fired emission control system was designed by B&W as an integral part of the steam generation plant. It was designed to treat the boiler flue gas flow of 78,000 acmm (2,760,000 acfm) (9,800 acmm or 345,000 acfm per module at 140°C or 285°F). The ductwork design does not provide for flue gas bypass of the system. Also, the plant does not have an alternate or secondary fuel supply. Each module can be isolated for maintenance by individual dampers. On site limestone grinding and slurry storage facilities provide up to 907 Mg (1,000 tons) of slurry per hour. The unit has a balanced draft system with three 7,000 hp forced draft fans and six 7,000 hp induced draft fans located between the emission control system and the 213 m (700 foot) stack. There is a common plenum at both the scrubber inlet and outlet. Spent slurry and fly ash are removed from the module recirculation tank through rubber lined pipes to a settling pond at the rate of 3,175 Mg (3,500 tons) of solids per day. Clear make-up water is pumped from the pond and the loop is closed by recycling ball mill and module make-up water back into the system.²

TABLE 3.
BOILER NO. 5 DESIGN DATA

Type	Oil/Gas
Manufacturer	Babcock and Wilcox
Number of Burners	24
Burner Arrangement	Front and Rear Firing
Air Preheater	Yes
Combustion Air Temperature	299°C (570°F)
Combustion Air Volume	20,200 Nm ³ /min/27°C (766,000 scfm/80°F)
Recirculation Gas	Yes
Volume	10,900 Nm ³ /min (414,000 scfm)
Temperature	374°C (705°F)
Reheat	Two Stage
Design Steam Rate	
Super Heat	970,000 kg/hr, 25 MPa, 538°C (2,135,000 lb/hr, 3650 psig at 1000°F)
First Reheat	821,000 kg/hr, 7.3 MPa, 562°C (1,807,000 lb/hr, 1065 psig at 1025°F)
Second Reheat	726,000 kg/hr, 2.6 MPa, 566°C (1,598,000 lb/hr, 365 psig at 1050°F)

In abbreviated terms, as the hot flue gas enters the venturi, Figure 5, it is sprayed with slurry from 48 spray and 32 wall wash nozzles, resulting in up to 99 percent of the particles being agglomerated to the sump below. The gas continues through the sump making a 180 degree turn up through the absorber section. In the reaction chamber, the sulfur dioxide is removed as the gas is forced through a limestone slurry solution sprayed on stainless steel sieve trays. The chemical reaction in part combines the calcium carbonate, water and sulfur dioxide to form two relatively insoluble calcium salts in the sump, calcium sulfate and calcium sulfite. The cleaned gas passes through mist eliminators to remove moisture and then is reheated to avoid deposits on the fans and provide buoyancy from the stack.²

The oil fired utility boiler is designed to control the emission of sulfur dioxide by burning low sulfur content fuels. The emissions of particles and hydrocarbons are controlled by furnace design, operation, and maintenance to ensure complete combustion. Sub-stoichiometric combustion and flue gas recirculation are used to control the emission of nitrogen oxides. Sub-stoichiometric combustion involves two stages. The first stage is carried out in a fuel rich, air lean environment. The second stage uses air injected into the combustion zone to ensure complete combustion of the fuel. These techniques lower the flame temperature which results in reduced formation of nitrogen oxides.

Test Description and Conditions

Multimedia emission tests were conducted at the coal burning utility from April 8 through 19, 1978, and at the oil burning utility from August 30, 1978, through September 11, 1978. Gaseous, liquid, and solid emissions were sampled during coal and oil firing to obtain data for the comparative environmental impact assessment. Flue gas sampling was conducted before and after the scrubber at the coal fired utility to determine which pollutants are removed or modified by the control device.

Emissions were characterized using EPA's phased approach. This approach utilizes two levels of sampling and analysis (Level 1 and Level 2). Level 1 screening procedures are accurate within a factor of 2 to 3. They provide preliminary assessment data and identify problem areas and information gaps. Based on these data, a site specific Level 2 sampling and analysis plan is developed. Level 2 provides more accurate and detailed information to confirm and expand on the information gathered in Level 1. The methods and procedures used for Level 1 are documented in the manual, "Combustion Source Assessment Methods and Procedures Manual for Sampling and Analysis," September 1977. The Level 2 methods and procedures include "state-of-the art" techniques required for this particular site.³

Normally all Level 1 samples are analyzed and evaluated before moving to Level 2. Because of the program time constraints, the Level 1 and Level 2 samples were obtained during the same test period.

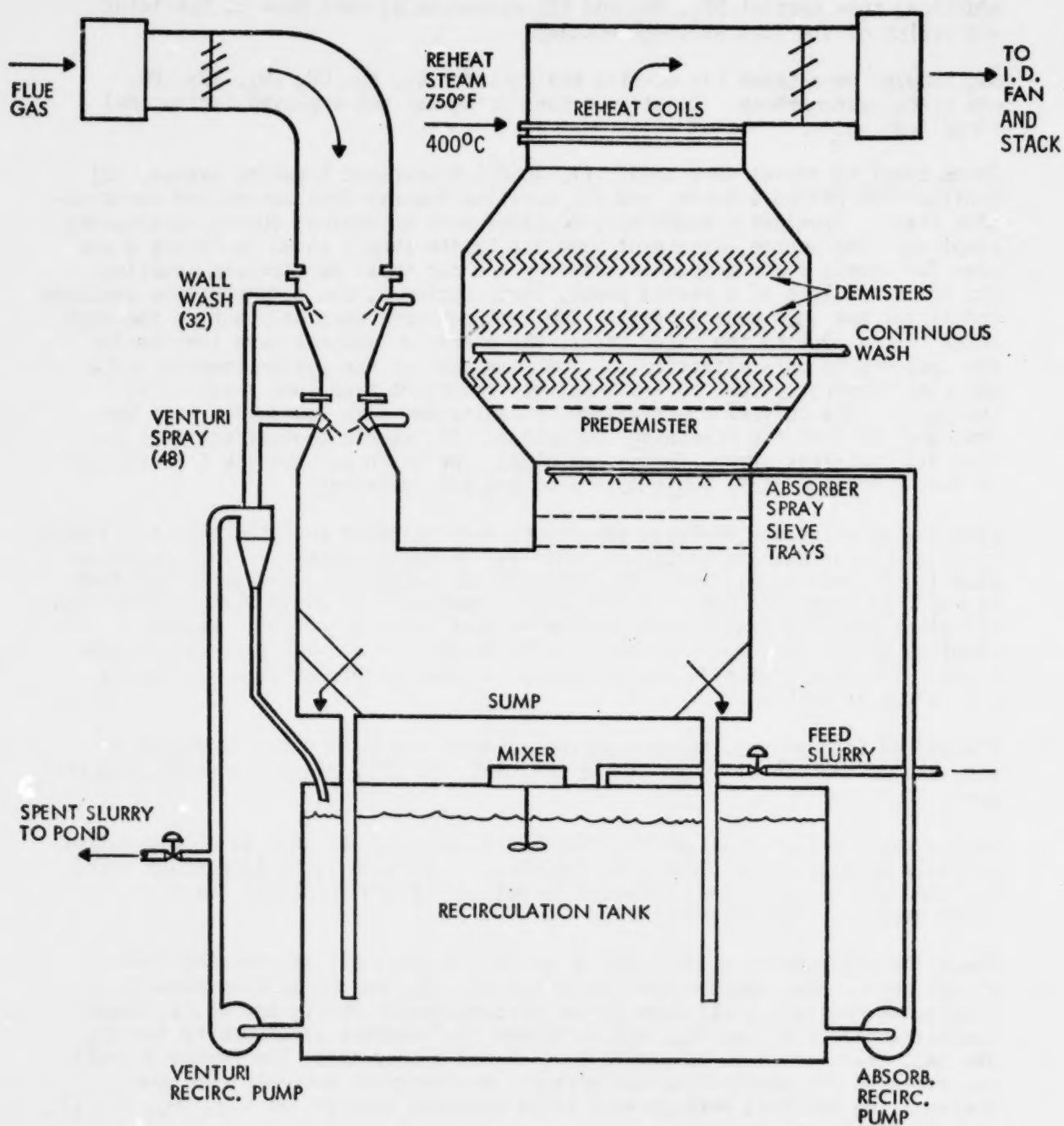


FIGURE 5
COAL FIRED BOILER FGD MODULE

Gaseous Effluents

The boiler flue gas at the coal burning utility was sampled at the FGD inlet gas manifold before the gas flow was divided to the eight scrubber modules. Comparative scrubber outlet data were obtained by sampling at the stack. In addition, some special SO_2 , SO_3 and SO_4 measurements were made at the inlet and outlet of the last scrubber module.

Bag samples were taken for on-site analysis for O_2 , N_2 , CO , CO_2 , SO_2 , NO , and C_1 - C_6 hydrocarbons. Continuous monitoring was not employed during coal fired testing.

Three sampling trains were used: (1) Source Assessment Sampling System, (2) modified EPA Method 5 train, and (3) modified Goksoyr-Ross controlled condensation train. Isokinetic sampling conditions were maintained during particulate sampling. The Source Assessment Sampling System (SASS) shown in Figure 6 was used for Levels 1 and 2 organic sampling and for total particulate sampling. The train consists of a heated probe, three cyclones, and a filter (the cyclones and filter are in a heated oven). The cyclones were used only during the coal inlet tests. During the other tests, the particle loadings were too low for the cyclones to work effectively. The remainder of the system consists of a gas conditioning system, an XAD-2 polymer absorbent trap, and a series of impingers. The polymer traps gaseous organics and some inorganics, and the impingers collect the remaining inorganics. All sample contact surfaces are Type 316 stainless steel, Teflon, or glass. The train was run for 6 to 8 hours so that a minimum of 30 cubic meters of gas was collected.

Previous sampling and analysis experience had indicated that the SASS train materials might contaminate certain organic and inorganic samples. The contamination is of concern only when the pollutant is present at a concentration that is near the detection limit of the Level 2 methods. To avoid that possibility, all-glass sampling trains were used to collect Level 2 samples. Method 5 sampling trains were modified as shown in Figure 7 for total particulate sampling. This train sampled approximately 10 cubic meters of flue gas during a 6 to 9 hour test run.

A modified Goksoyr-Ross controlled condensation train as shown in Figure 8 was used at each location to sample SO_3 , SO_4 , F^- , Cl^- , HF , HCl , and SO_2 (impinger).

During Level 2 test runs, an MRI impactor (Figure 9) was used to obtain outlet particle samples by particle size fraction. Polarized light microscopy (PLM) was used on the particles collected by Method 5 train to obtain the inlet particle size distribution.

Since the oil burning utility had no control device, all gas was sampled at the stack. Bag samples were taken for N_2 , CO_2 , and C_1 - C_6 hydrocarbons. Continuous monitoring was done by gas chromatography for O_2 and CO , by chemiluminescence for NO and NO_x , and by pulsed fluorescence spectrography for SO_2 . The SASS train was used to sample Level 1 and 2 organics. The Method 5 train was employed for particulate collection. No inorganic analyses were performed. The modified Goksoyr-Ross train provided samples for SO_3 , SO_4 , F^- , Cl^- , and SO_2 (impinger) analyses.

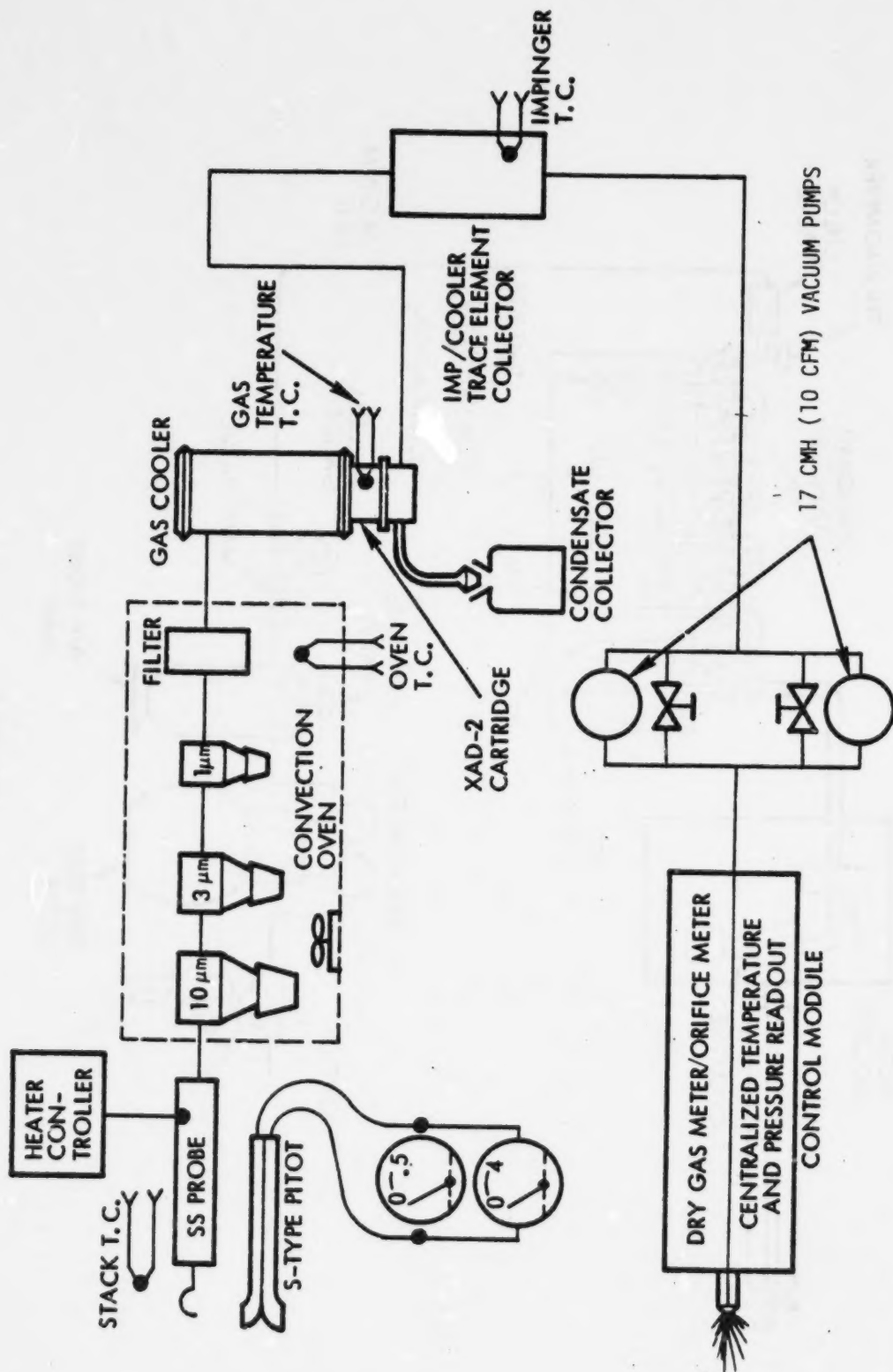


FIGURE 6
SOURCE ASSESSMENT SAMPLING SYSTEM (SASS) SCHEMATIC

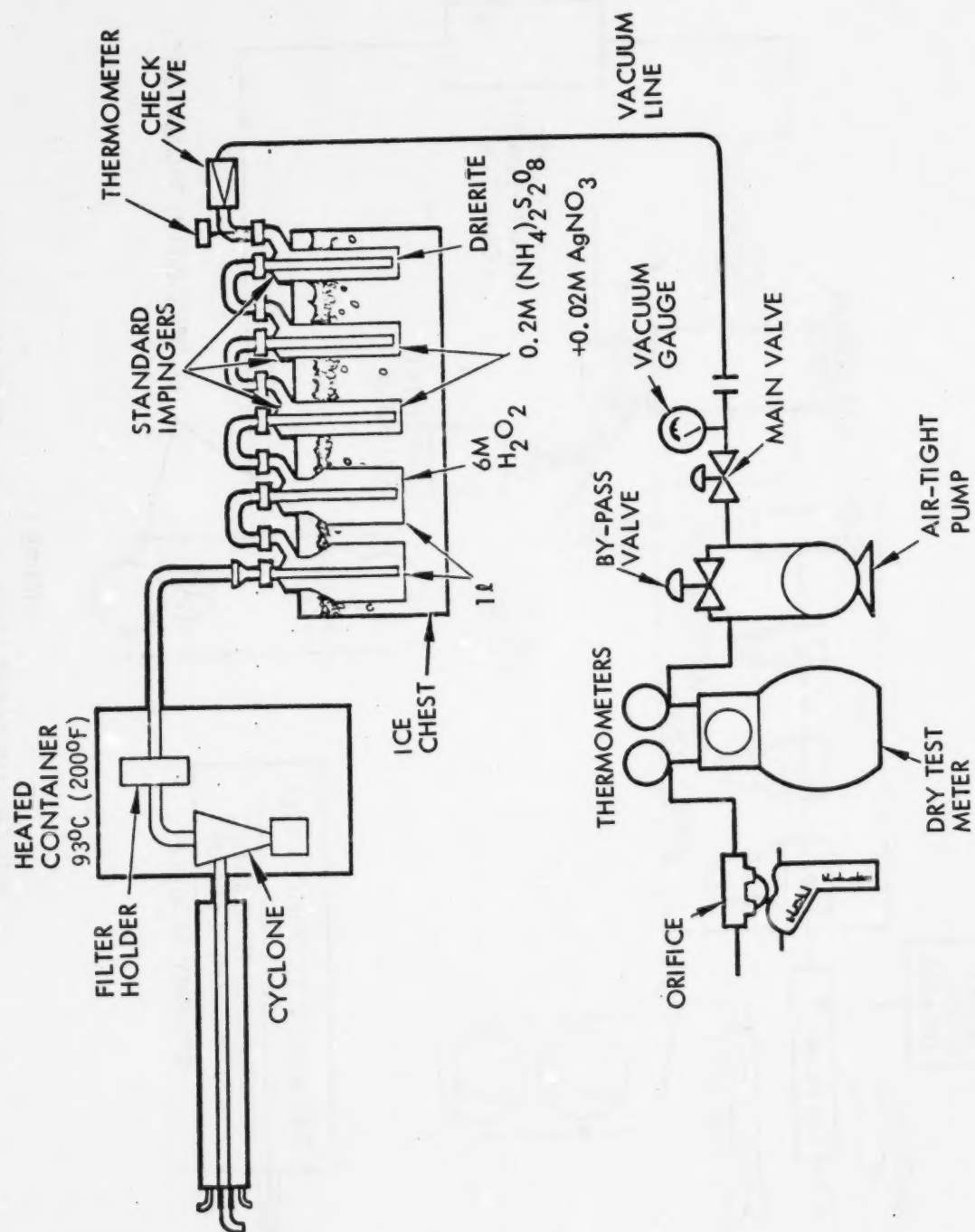


FIGURE 7
MODIFIED METHOD 5 SAMPLING SYSTEM SCHEMATIC

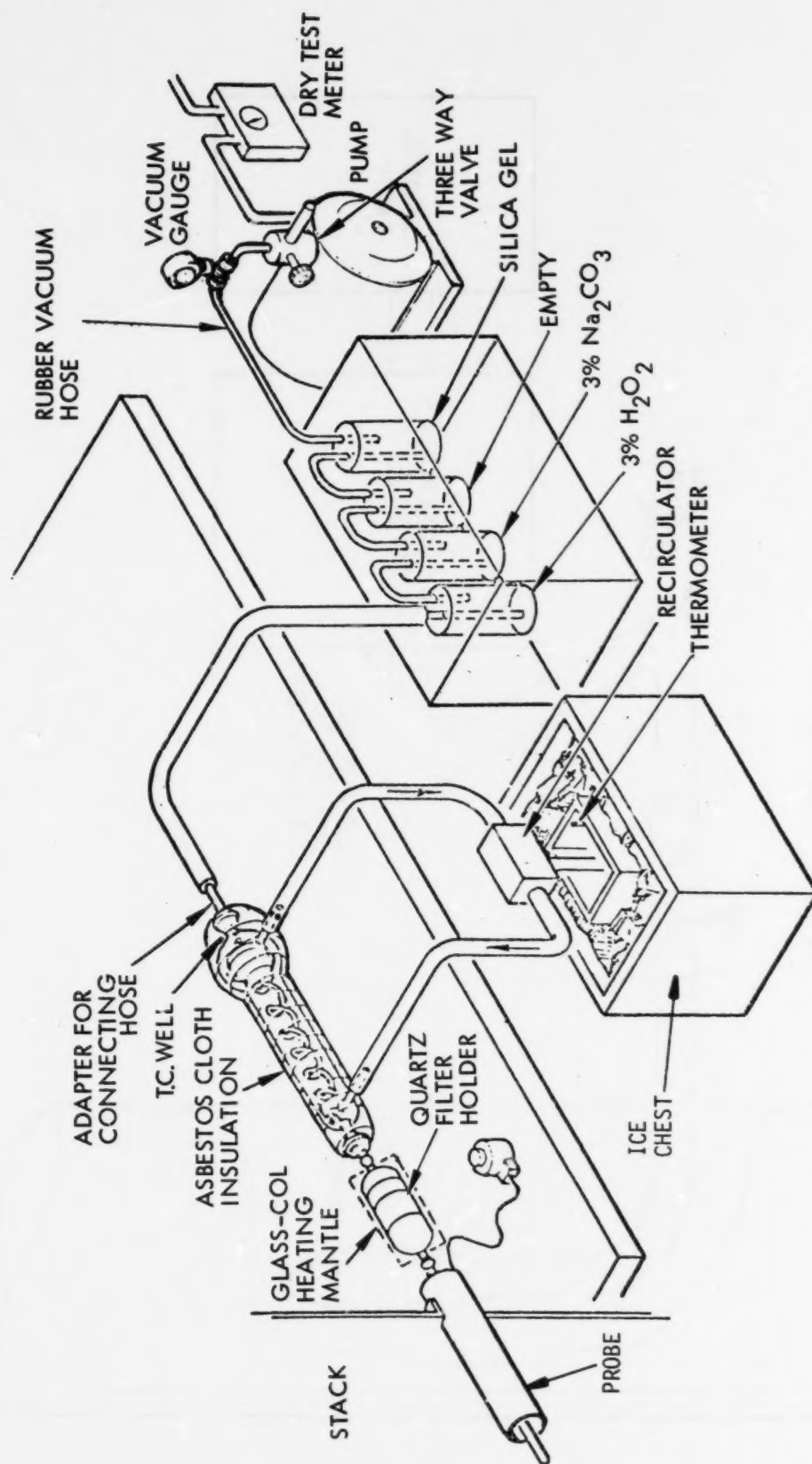


FIGURE 8
 CONTROLLED CONDENSATION SAMPLING SYSTEM SCHEMATIC

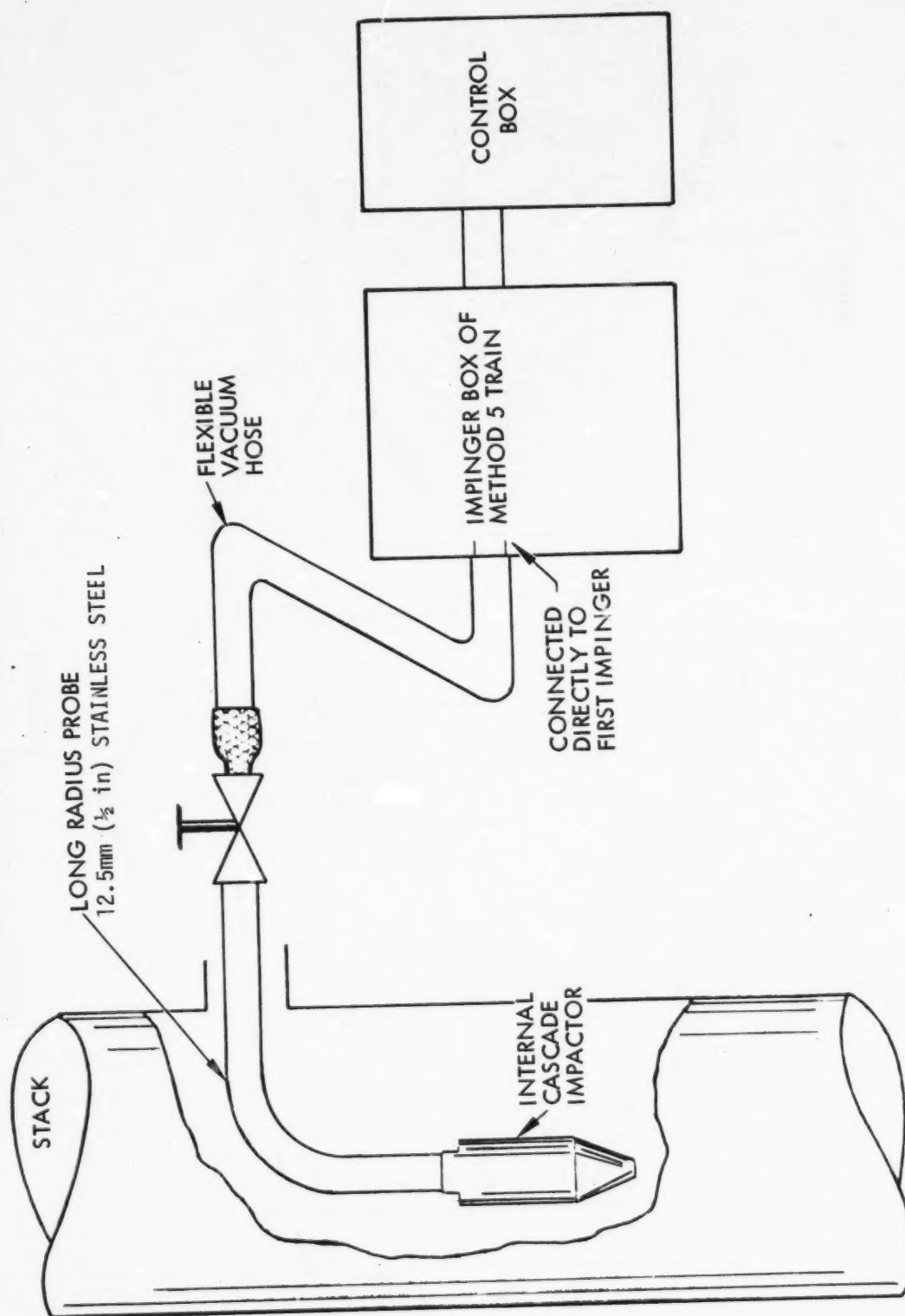


FIGURE 9
MRI IMPACTOR SCHEMATIC

Liquid Effluents

The liquid effluent from the scrubbing system installed at the coal burning utility is a side stream of the slurry recycling system. Approximately 1,500 l/min (400 gpm) of spent slurry is discharged from each scrubber module (a total of 12,100 l/min or 3,200 gpm for the entire scrubbing system) and piped to a settling pond on site. Table 4 gives the approximate analysis of the spent slurry solution.² Upon entering the settling pond, the slurry is diluted and the suspended solids settle. The clarified solution is recycled to the scrubber or discharged to a cooling lake.

TABLE 4.

COMPOSITION OF SPENT SCRUBBER SLURRY

CaCO ₃	46 g/l
CaSO ₃	50 g/l
CaSO ₄	16 g/l
Flyash	30 g/l
Total Solids	14%
pH	5.6

There is no liquid effluent connected with air quality control from the oil burning utility.

Solid Effluent

At the coal burning plant, the scrubber solids are discharged to a settling pond at a rate of approximately 132 Mg/hr. The majority of this waste material is disposed of in a company owned landfill.

Dredged solids from the settling basins at the oil fired utility are hauled to municipal landfills. No solids are discharged directly to the environment.

Laboratory Analyses

The samples from the various sampling trains were returned to the laboratory for analysis. Detailed analysis procedures can be found in the manual "Combustion Source Assessment Methods and Procedures Manual for Sampling and Analysis," September 1977.

Level 1 analyses for particles and gases were made for inorganics by SSMS and for selected anions and organics by liquid chromatography, infrared, and mass spectroscopy. Solids, slurries, and liquids were similarly analyzed, although the work-up procedures were different.

More detailed and more quantitative Level 2 analyses were performed to identify and quantitate specific compounds indicated by the Level 1 analyses.

Test Conditions

Five tests were performed with the coal burning utility boiler. Limited supplementary coal firing tests were performed using the controlled condensation train only. Four tests were performed with the oil burning utility boiler. Test conditions are summarized in Table 5.

The flue gas entered the scrubber at the coal burning utility at about 163°C (325°F) and exited at about 93°C (200°F). The stack gas at the oil burning utility was about 121°C (250°F).

The coal unit operated at 71 to 87 percent of design; the oil unit, at 62 to 94 percent.

MULTIMEDIA EMISSION RATES

Gaseous Emissions

As discussed previously, flue gas generated by the furnaces at the coal fired power plant were passed through a two stage venturi absorber limestone slurry scrubbing system prior to release to the atmosphere through a tall stack. The oil fired boiler tested, on the other hand, operated with flue gas recirculation for NO_x control but was not equipped with an SO₂ removal system. Emissions information presented in the following sections for the coal fired plant normally includes both scrubber inlet and scrubber outlet (stack) data, whereas information presented for the oil-fired plant is for stack emissions.

Criteria Pollutants

Federal New Source Performance Standards (NSPS) currently in effect define allowable emission rates of NO_x (as NO₂), SO₂, and total particles from fossil fuel fired utility boilers having 25 MW output or greater. More stringent limitations have been proposed by EPA for NO_x, SO₂, and total particles emissions. Federal NSPS do not currently address either CO or total hydrocarbon emissions. Existing NSPS and corresponding proposed or potential emission standards for coal- and oil-fired utility boilers are summarized in Table 6.

TABLE 5
SUMMARY OF TEST CONDITIONS

Test No.	Electrical Output (Gross-MW)	% of Maximum Boiler Load ^a	Nominal Fuel Feed Rate kg/hr	% O ₂ in Flue Gas	Excess Air at Furnace ^b
COAL FIRING					
132	620	71	254,000	6.0	~ 15
133	640	73	254,000	6.0	~ 15
134	690	79	295,000	6.0	~ 15
135	760	87	318,000	6.0	~ 15
136	760	87	300,000	6.0	~ 15
OIL FIRING					
141	330	94.3	69,008	6.35	~ 20
142	218	62.3	54,934	5.76	~ 20
143	325	92.9	66,284	5.66	~ 20
144	310	85.7	61,744	5.98	~ 20

^aBased on gross megawatts produced.

^bFull load excess air level. At lower loads, slightly higher levels are expected. No allowance made for inleakage.

TABLE 6.
EXISTING AND PROPOSED FEDERAL EMISSION STANDARDS
ng/J (1b/10⁶ Btu)

	Coal Fired Utilities		Oil Fired Utilities	
	NSPS	Proposed Standard	NSPS	Proposed Standard
NO _x (as NO ₂)	300 (0.70)	260 (0.60)	130 (0.30)	130 (0.30)
SO ₂	520 (1.20)	520 (1.20) max. with 85% reduction to 85 (0.20)	344 (0.80)	520 (1.20) max. with 85% reduction to 85 (0.20)
Total Particles	43 (0.10)	13 (0.03) max. with 99% reduction	43 (0.10)	13 (0.03)

Five tests were performed to determine the emissions from the coal fired boiler. A summary of the criteria pollutant emissions data for the five test series is presented in Table 7. For the oil fired boiler, four tests were performed with oil firing. Criteria pollutant emissions data for these tests are summarized in Table 8. In Table 9, average emissions data from the coal fired and oil fired boilers are presented. The emissions data are discussed by specific pollutant in the ensuing subsections.

Particles

Total

Average emissions of total particles were 1090 ng/J for coal firing prior to scrubbing and 7.5 ng/J for oil firing. Total particles emissions from the coal fired boiler after scrubbing were 80 ng/J. This corresponds to 93 percent particle removal efficiency for the scrubber. Controlled particle emissions from the coal fired boiler tested are still above the NSPS limit of 43 ng/J for utility boilers. Particle emissions from the oil fired boiler tested, however, are well below this limit. When comparing stack emissions, release of particles to the atmosphere from coal firing is 10 times greater than from oil firing on an emission factor basis.

Size Distribution

For the coal fired boiler, size distributions of particles at the scrubber inlet and outlet were determined by two methods. Due to the high particle

TABLE 7. SUMMARY OF CRITERIA POLLUTANT EMISSIONS - COAL FIRING

Test No.	NO _x (as NO ₂)	CO ^a	SO ₂ ^b	Emission Factor (ng/J)				Total Organics	Total Organics	Total Particulates
				C ₁ -C ₆ Organics	C ₇ -C ₁₆ Organics	>C ₁₆ Organics				
132 Inlet	460	≤500	3210	ND ^c	ND	ND		ND	ND	ND
132 Outlet	300	≤500	640	ND	ND	ND		ND	ND	ND
133 Inlet	250	≤500	3440	ND	ND	ND		ND	ND	ND
133 Outlet	190	≤500	820	ND	ND	ND		ND	ND	ND
134 Inlet	480	≤500	2970 ^d	1-2.3	0.45	1.32		2.77-4.07	1280	
134 Outlet	380	≤500	240	0.85-2.0	0.12	0.48		1.45-2.60	ND	
135 Inlet	700	≤550	3650	ND	ND	ND		ND	900	
135 Outlet	370	≤550	ND	ND	ND	ND		ND	80	
136 Inlet	730	≤550	2560	ND	ND	ND		ND	ND	
136 Outlet	400	≤550	770	ND	ND	ND		ND	ND	
Average Inlet	520	≤520	3170	1-2.3	0.45	1.32		2.77-4.07	1090	
Average Outlet	330	≤520	740	0.85-2.0	0.12	0.48		1.45-2.60	80	

^a CO emission factor was based on the detection limit of 1000 ppm.

^b SO₂ emissions were determined from grab bag sampling for Test Nos. 132-134, and from the impinger solution of the controlled condensation system for Test Nos. 135 and 136.

^c ND - data not available.

^d Determination of SO₂ emissions at the scrubber outlet for Test No. 134 appears to be in error. This data point was judged to be an outlier at 90% probability level by the method of Dixon, and discarded in the computation of average SO₂ emissions.

TABLE 8. SUMMARY OF CRITERIA POLLUTANT EMISSIONS - OIL FIRING

Test No.	NO _x (as NO ₂)	CO	SO ₂	Emission Factor (ng/J)				Total Organics	Total Organics	Total Particulates
				C ₁ -C ₆ Organics	C ₇ -C ₁₆ Organics	>C ₁₆ Organics				
141	114	11.3	105	0.35-0.56	ND ^a	ND		0.51-0.74 ^b	ND	ND
142	91	5.2	95	0.29-0.43	0.02	0.16		0.48-0.61	8.3	8.3
143	117	5.6	105	0.11-0.22	0.02	0.14		0.27-0.38	6.6	6.6
144	98	4.5	103	ND	ND	ND		ND	ND	ND
Average	105	6.6	102	0.25-0.40	0.02	0.15		0.42-0.58	7.5	7.5

^a ND - data not available.^b Total organics for Test No. 141 were estimated using C₇-C₁₆ and >C₁₆ organics emissions from Test Nos. 142 and 143.

TABLE 9. CRITERIA POLLUTANT EMISSIONS FROM UTILITY BOILERS

Pollutant	Emission Factor (ng/J)		
	Coal Firing		Oil Firing
	Before Scrubber	After Scrubber	
NO _x (as NO ₂ near full load)	715 ^a	385 ^a	116
CO	≤520	≤520	6.6
SO ₂	3170	740	102
Total Organics	2.77-4.07	1.45-2.60	0.42-0.58
Total Particulates	1090	80	7.5

^aComputed using average NO_x emissions from Test Nos. 135 and 136. These two tests were conducted at 87% boiler load.

loading at the scrubber inlet during coal firing, polarized light microscope (PLM) analyses were utilized to obtain a size distribution in terms of optical diameter and number of particles per size range. All other particle size distribution determinations involved streams with substantially lower solids loadings; therefore, an MRI cascade impactor was used. The cascade impactor data differs from PLM analyses in that size distributions are determined in terms of aerodynamic diameter and weight percent in each size range. Thus, data from the two methods cannot be directly compared. For this reason, the PLM data have been converted to the same basis as the impactor data by assuming that particle density is independent of particle diameter. This is a reasonable assumption because the major components of the particles generated from coal combustion, the aluminosilicates and iron oxides, are known to partition equally between small and large particles. With the constant density assumption, the weight distribution in each size range would be proportional to the product of the number distribution and the particle volume representing the size range. The particle volume was calculated based on the geometric mean diameter for the size range.

Particle size distribution data from coal firing tests are summarized in Table 10. These data show a significant change in particle size distribution before and after scrubbing. The increase in the fraction of fine particles across the scrubber indicates that coarse particles were removed more efficiently than fine particles. Particles larger than $3\text{ }\mu\text{m}$ were removed with efficiencies of greater than 99 percent while particles smaller than $3\text{ }\mu\text{m}$ actually showed a net increase in emission rates across the scrubber. This net increase indicates that the venturi scrubber at the coal fired plant is probably not effective in removing fine particles present in the flue gas. Additionally, the net increase also raises the possibility that fine particles may be generated within the scrubber, or that the particle size distribution may be modified during the high energy scrubbing process.

Particle size distribution data are not available from the oil firing tests. However, data available from the literature have indicated that emitted particles are generally 90 wt percent less than $7\text{ }\mu\text{m}$ for oil fired boilers.⁵

In a recent paper from EPA's Health Effects Research Laboratory,⁶ it is stated that larger particles (from 3 to $15\text{ }\mu\text{m}$) deposited in the upper respiratory system (in the nasopharynx and conducting airways) can be associated with health problems. This is in contrast to the past belief that particles of health consequence were those less than $3\text{ }\mu\text{m}$ size and deep lung penetrable. The area of hazard now is particles which are $15\text{ }\mu\text{m}$ and less, which have been designated as "inhaled particles" (IP).

For oil firing, it can be reasonably assumed that almost all the particles emitted are $15\text{ }\mu\text{m}$ or less. Emissions of inhalable particles at the coal fired plant after scrubbing were approximately 75 ng/J , as compared to 7.5 ng/J from the oil fired plant. Again, emissions of inhalable particles from coal firing are 10 times greater than from oil firing, even after control.

TABLE 10. SCRUBBER INLET AND OUTLET PARTICULATE SIZE DISTRIBUTION - COAL FIRING

Aerodynamic Diameter Size Range, μm	Weight %		Emission Factor (ng/J)		Removal Efficiency, %
	Scrubber Inlet	Scrubber Outlet	Scrubber Inlet	Scrubber Outlet	
< 1	<0.01	82	<0.1	65.6	<0
1 - 3	<0.3	11	<3	8.8	<0
3 - 10	13	1	141	0.8	99.4
> 10	87	6	946	4.8	99.5
Total	100	100	1090	80	92.7

Sulfur Compounds

For the coal fired boiler, SO_2 emissions were determined from grab bag sampling by pulsed fluorescent analyzer and from the impinger solution of the controlled condensation system by titration. For the oil fired boiler, SO_2 emissions were monitored continuously by pulsed fluorescent analyzer.

Average SO_2 emissions were 3170 ng/J for coal firing prior to scrubbing and 102 ng/J for oil firing. Average SO_2 emission rates from the coal fired boiler were 740 ng/J after scrubbing. This represents a mean scrubber efficiency of 77 percent for SO_2 removal, within the design range of 75-80 percent. For the oil fired boiler tested, the measured SO_2 emissions of 102 ng/J are well below the NSPS limit of 344 ng/J for oil fired utility boilers.

In addition to the grab bag sampling at the coal fired plant and continuous monitoring at the oil fired plant to determine SO_2 in flue gas emissions, the modified Goksoyr-Ross controlled condensation system was also used at both plant locations to determine SO_2 , SO_3 , and SO_4 emissions. A summary of these sulfur compound determinations is presented in Table 11. Based on the total sulfur compounds present in the flue gas, an average of 98.5 percent of the input sulfur is emitted as SO_2 in coal firing when emissions are uncontrolled, and 97.7 percent of the input sulfur is emitted as SO_2 in oil firing. The removal efficiency for SO_2 by the scrubber system at the coal fired plant averaged 77 percent as previously discussed. About 80 percent of the SO_3 , based on average inlet and outlet concentrations, was also removed by the scrubber. Removal efficiency for SO_4 averaged 83 percent, again based on averages, which is lower than the total particle removal efficiency. This is an indication that a large fraction of SO_4 may be associated with the fine particles in the flue gas stream from the coal fired boiler, which are less efficiently scrubbed.

A comparison of the sulfur oxide emissions is best made by comparing emission factors which are expressed as mole percent of total sulfur compounds (SO_2 , SO_3 , and SO_4) present in the flue gas. The percentages of total sulfur compounds present as SO_2 are almost identical for coal and oil firing. For SO_3 , the values were an average of 1.1 percent for coal firing and 1.2 percent for oil firing. The SO_3 emissions during oil firing were lower than typical values reported in the literature, probably because of the lower vanadium and nickel content of the fuel oil burned, and the use of flue gas recirculation which reduces the oxygen concentration available for SO_3 formation. Conversion of sulfur compounds to SO_4 was an average of 0.4 percent for coal firing and 1.1 percent for oil firing. However, actual sulfate emissions were higher for coal firing due to the higher fuel sulfur content of coal.

Nitrogen Oxides

NO_x emissions from the coal fired boiler were determined by chemiluminescence from grab bag sampling. For the oil fired boiler, NO_x emissions were monitored continuously by chemiluminescent instrumentation. The NO_x emission factors

TABLE 11. SO₂, SO₃, AND SO₄⁼ EMISSIONS FROM COAL AND OIL FIRING

Sulfur Compound	Coal Firing				Oil Firing			
	Test No.	Scrubber Inlet ng/J	Scrubber Outlet ng/J	Mole % of Total Sulfur Compounds in Flue Gas	Removal Efficiency %	Test No.	Flue Gas ng/J	Mole % of Total Sulfur Compounds in Flue Gas
SO ₂	135	3650	ND ^a	97.6	--	143	75.5	97.7
	136	2560	770	97.8	70.0			
	I	3260	ND	98.0	--			
	II	3450	ND	98.7	--			
	III	3980	ND	99.1	--			
	IV	3640	ND	98.9	--			
	V	3610	ND	98.8	--			
SO ₃	135	99	17	2.1	82.8	143	1.1	1.2
	136	54	2	1.7	96.3			
	I	49	ND	1.2	--			
	II	39	ND	0.9	--			
	III	24	ND	0.5	--			
	IV	35	ND	0.8	--			
	V	39	ND	0.9	--			
SO ₄ ⁼	135	15.5	5.1	0.3	67.1	143	1.3	1.1
	136	18.2	2.4	0.5	86.8			
	I	41.2	ND	0.8	--			
	II	22.2	ND	0.4	--			
	III	24.2	ND	0.4	--			
	IV	18.6	ND	0.3	--			
	V	15.3	ND	0.3	--			

^a ND - data not available.

near full load conditions were 715 ng/J for coal firing prior to scrubbing and 116 ng/J for oil firing. Examination of the published data on NO_x emissions from coal fired cyclone boilers indicates an average emission factor of 662 ng/J.⁸ Thus, the NO_x emission factor for the coal fired boiler tested is well within the range typical of coal fired cyclone boilers. For oil firing, the lower NO_x emission factor was due to the use of flue gas recirculation for NO_x control.

NO_x data from the coal fired boiler tested generally indicate a significant reduction of NO_x across the scrubber. However, these NO_x data were determined by chemiluminescence from grab bag sampling, and it has been recently established that NO_x decay inside these grab bags is rapid in the presence of air.⁹ NO_x removal by the FGD system at the coal fired power plant was therefore probably not a real phenomenon. Additional data from on line monitoring are needed to confirm any NO_x reduction across the scrubber.

The measured NO_x emissions from the coal fired boiler tested at both the scrubber inlet and scrubber outlet exceed the NSPS limit of 300 ng/J for coal fired utility boilers. On the other hand, the NO_x emissions from the oil fired boiler tested at the stack are slightly below the NSPS limit of 130 ng/J for oil fired utility boilers. NO_x emissions from the coal fired boiler (at the scrubber inlet) are approximately 6 times those from the oil fired boiler. The differences in NO_x emissions are attributed to three factors: (1) higher thermal NO_x generation in cyclone furnaces because of extremely high heat release rates and the resulting high furnace gas temperatures; (2) higher fuel nitrogen content in coal as compared to oil, leading to higher fuel NO_x generation in the coal fired boiler; and (3) the use of flue gas recirculation for NO_x control at the oil fired boiler tested.

Carbon Monoxide

Average CO emissions from the oil fired boiler tested were 6.6 ng/J. No effort was made for accurate determination of CO from the coal fired boiler. The reported CO emission factor of less than 520 ng/J for the coal fired boiler was based on detection limits of the instrumentation. However, CO emissions from coal fired and oil fired utility boilers should be comparable in magnitude.

Organics

In the determination of organic emissions, gas chromatographic analyses were performed on grab bag samples of flue gas and catches from the Level 1 sampling (SASS train). Additionally, gravimetric analyses were performed on Level 1 samples to quantify high molecular weight organics. Each bag sample was collected over an interval of 30 to 45 minutes, with a single sample being collected per test. These samples were utilized to measure C_1 to C_6 hydrocarbons. The SASS train collects approximately 30 cubic meters of flue gas isokinetically during the test. Samples from the SASS train

were analyzed to determine organics higher than C₆. The C₇ to C₁₆ fraction was determined by gas chromatograph while organics higher than C₁₆ were determined gravimetrically.

Average organic emissions were 2.8-4.1 ng/J for coal firing prior to scrubbing and 0.4-0.6 ng/J for oil firing. Comparison of the data presented indicates that emissions from coal firing are greater than those from oil firing in C₁-C₆ organics, C₇-C₁₆ organics, and high molecular weight (>C₁₆) organics. At the coal fired plant, emissions of organics after scrubbing were 0.85-2.0 ng/J for the C₁-C₆ fraction (14 percent removal), 0.12 ng/J for the C₇-C₁₆ fraction (72 percent removal), and 0.48 ng/J for the high molecular weight fraction (63 percent removal). However, the organic emissions from coal firing after scrubbing are still greater than those from oil firing in all fractions.

Selected samples from the coal firing and oil firing tests were analyzed by combined gas chromatography/mass spectrometry (GC/MS) for the identification of organic compounds present. For the coal fired boiler, the organic compounds identified include aliphatic hydrocarbons, substituted benzene, ethylbenzaldehyde, dimethylbenzaldehyde, 2,6-pereriden-dione-4, and 2,6 dimethyl-2,5-heptadion-4-one, and the methyl ester of a long chain acid, at concentration levels ranging from 0.2 to 20 µg/m³ in the flue gas prior to scrubbing. With the exception of ethylbenzaldehyde, substituted benzenes, and aliphatic hydrocarbons, none of the other organic compounds were identified at the scrubber outlet. For the oil fired boiler, the organic compounds identified include aliphatic hydrocarbons, benzaldehyde, trimethyl cyclohexane-one, C₂ substituted acetophenone, the methyl esters of benzoic acid and substituted benzoic acid, diethylphthalate, and ethylbenzaldehyde, at concentration levels between 0.02 and 2 µg/m³. Thus, the organic compounds from coal firing and oil firing are somewhat similar.

Emissions of polycyclic organic matter (POM) determined by GC/MS for coal firing and oil firing are summarized in Tables 12 and 13. For coal firing, most of the POM compounds identified are naphthalene, substituted naphthalenes, biphenyl, and substituted biphenyls. No POM compounds were identified at the scrubber outlet of the coal fired plant. POM compounds found at the scrubber inlet are at levels several orders of magnitude below their respective MATE values. The Minimum Acute Toxicity Effluent (MATE) values are emission level goals developed under direction of EPA, and can be considered as concentrations of pollutants in undiluted emission streams that will not adversely affect persons or ecological systems exposed for short periods of time.¹⁰

The total POM emissions from oil firing are lower than those from coal firing. Again, naphthalene is the principal component of the POM emissions. With the exception of the possible presence of benzo(a)pyrene, all POM compounds from oil firing are at levels too low to be of environmental concern.

Inorganics

For the coal fired boiler, trace elements present in the flue gas were determined using atomic absorption spectroscopy (AAS). For the oil fired boiler,

TABLE 12. POM EMISSIONS FROM COAL FIRING
PRIOR TO SCRUBBING

Compound	Emission Concentration $\mu\text{g}/\text{m}^3$	MATE Value $\mu\text{g}/\text{m}^3$	Potential Degree of Hazard ^e
Decahydronaphthalene ^a	0.1	130,000	<0.0001
Ditert-butyl naphthalene ^b	0.3	230,000	<0.0001
Dimethyl isopropyl naphthalene ^b	0.3	230,000	<0.0001
Hexamethyl biphenyl ^c	0.6	1,000	0.0006
Hexamethyl, hexahydro indacene	1.0	No data	--
Dihydronaphthalene ^a	0.03	130,000	<0.0001
C ₁₀ substituted naphthalene ^b	0.06	230,000	<0.0001
C ₁₀ substituted decahydro- naphthalene ^a	1.0	130,000	<0.0001
Methyl naphthalene	1.6	130,000	<0.0001
Anthracene/phenanthrene	0.3	1,600	0.0002
1-1'-biphenyl	4.0	1,000	0.004
9,10-dihydronaphthalene ^a / 1-1'-diphenylethene	0.2	130,000	<0.0001
1,1-bis(p-ethyl phenyl)-ethane/ tetramethyl biphenyl ^a	9.0	1,000	0.009
5-methyl-benz-c-acridine ^d	0.2	10,500	<0.0001
2,3 dimethyl decahydro- naphthalene	<0.03	130,000	<0.0001
Total	18.7		

^a The MATE values for decahydronaphthalene, dihydronaphthalene and any substituted decahydronaphthalene are assumed to be the same as that for tetrahydronaphthalene.

^b The MATE values for alkyl naphthalenes are assumed to be the same as that for methyl naphthalene.

^c The MATE value for hexamethyl biphenyl is assumed to be the same as that for biphenyl.

^d The MATE value for 5-methyl-benz-c-acridine is assumed to be the same as that for benz(c)acridine.

^e Calculation carried to four significant figures.

TABLE 13. POM EMISSIONS FROM OIL FIRING

Compound	Emission Concentration, $\mu\text{g}/\text{m}^3$		MATE Value $\mu\text{g}/\text{m}^3$	Potential Degree of Hazard	
	Test 142	Test 143		Test 142	Test 143
Naphthalene	7	3	50,000	0.0001	<0.0001
Phenanthridine ^a	0.3	--	No data	ND	--
Dibenzothiophene ^b	0.6	--	23,000	<0.0001	--
Anthracene/ phenanthrene	1	0.2	1,600	0.0006	0.0001
Fluoranthene	1	--	90,000	<0.0001	--
Pyrene	1	--	230,000	<0.0001	--
Chrysene/ benz(a)anthracene	0.1	--	45	0.0022	--
Benzopyrene ^c and perylene	0.04	--	0.02	2	--
Tetramethyl- phenanthrene ^d	--	0.6	1,600	--	0.0004
Total	11.0	3.8			

^a The presence of phenanthridine has not been positively identified.

^b The presence of dibenzothiophene has not been positively identified. Also, the MATE value for dibenzothiophene is assumed to be the same as that for benzothiophene.

^c The MATE value for benzo(a)pyrene is used in the computation of the potential degree of hazard.

^d The MATE value for tetramethyl phenanthrene is assumed to be the same as that for phenanthrene.

trace element concentrations in the flue gas were computed by assuming that all trace elements present in the fuel oil are emitted in the stack. Trace elements in the fuel oil were determined using spark source mass spectrometry (SSMS).

Concentrations of 18 major trace elements present in the flue gas during coal and oil firing are presented in Tables 14 and 15. To assess the hazard potential of these emissions, the emission concentrations are compared with the Minimum Acute Toxicity Effluent (MATE) values. The MATE values tabulated here represent air concentrations which were derived from human health considerations. Analysis of the flue gas generated during coal firing indicates that 16 elements exceeded their respective MATE values at the scrubber inlet and 7 exceeded their MATE values at the scrubber outlet. These seven elements which are of potential hazard are arsenic, cadmium, chromium, nickel, lead, iron, and zinc. During oil firing, only chromium and nickel exceeded their MATE values at the stack. The MATE values for nickel and chromium are extremely low due to considerations for potential human carcinogenicity. If Threshold Limit Values (TLV's) are used as the basis for comparison, then emissions of chromium and nickel from oil firing are respectively below and at their TLV's which are each 0.5 mg/m^3 .

Emission factors for the 18 trace elements analyzed are presented in Tables 16 and 17. Comparison of these emission factors shows that, with the exception of cobalt and nickel, emissions of trace elements from coal firing after scrubbing are considerably greater than corresponding uncontrolled emissions from oil firing. Also presented in Table 16 is the scrubber removal efficiency for each element during coal firing. An overall removal efficiency of 94 percent was obtained for these trace elements. However, some elements were removed with less than the average removal efficiency.

To better understand the removal efficiency of the individual trace elements, the enrichment factor has been computed for each trace element across the scrubber during coal firing. The enrichment factor is defined here as the ratio of the concentration of a trace element to that of aluminum in the scrubber outlet, divided by the corresponding ratio in the scrubber inlet. Aluminum is selected as the reference material because it has been known to partition equally among particles of different size.* The enrichment factors presented in Table 16 show that all the other 17 trace elements are enriched across the scrubber. The enrichment observed is due primarily to the partitioning of trace elements as a function of particle size, and the greater collection efficiency of the scrubber for the large size particles. Also note that many of the trace elements that show an enrichment trend, such as mercury, selenium and arsenic, either occur as element vapors or form volatile compounds at furnace temperatures. Condensation and surface adsorption of the more volatile elements or their oxides and halides downstream of the furnace could, therefore, result in higher concentrations of these trace elements on smaller particles.

* Silicon, iron, and scandium have also been used by other investigators as the reference element in the computation of enrichment factors. Notice that iron has an enrichment factor of 1.4 in this study while silicon and scandium were not measured.

TABLE 14. EMISSION CONCENTRATIONS OF TRACE ELEMENTS
DURING COAL FIRING

Element	Concentration, mg/m ³		MATE For Air (Health Basis), mg/m ³	Potential Degree ^a of Hazard	
	Scrubber Inlet	Scrubber Outlet		Scrubber Inlet	Scrubber Outlet
Al	132	3.0	5.2	25	0.58
As	0.98	0.94	0.002	490	470
Be	0.021	0.0018	0.002	11	0.90
Ca	49	2.0	16	3.1	0.13
Cd	5.1	0.58	0.010	520	58
Co	0.19	0.013	0.050	3.7	0.26
Cr	1.3	0.12	0.001	1300	120
Cu	1.2	0.19	0.20	6.0	0.95
Fe	401	13	1.0	400	13
Hg	0.095	0.0057	0.050	1.9	0.11
Mn	0.70	0.15	5.0	0.14	0.03
Ni	2.0	0.054	0.015	130	3.6
Pb	11	2.9	0.15	73	19
Sb	0.78	0.27	0.50	1.6	0.54
Se	0.37	0.088	0.20	1.8	0.44
Sr	0.46	0.038	3.0	0.15	0.013
V	0.78	0.083	0.50	1.6	0.17
Zn	105	21	4.0	26	5.2

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

TABLE 15. EMISSION CONCENTRATIONS OF TRACE ELEMENTS
DURING OIL FIRING

Element	Flue Gas Concentration mg/m ³	MATE for Air (Health Basis) mg/m ³	Potential Degree of Hazard ^a
Al ^b	0.084	5.2	0.016
As	<0.0007	0.002	<0.33
Be	<0.0007	0.002	<0.33
Ca ^b	1.66	16	0.10
Cd	<0.0007	0.010	<0.07
Co	0.03	0.050	0.6
Cr	0.006	0.001	6
Cu	0.007	0.20	0.035
Fe	0.1	1.0	0.1
Hg ^c	<0.005	0.050	<0.1
Mn	0.003	5.0	0.0006
Ni	0.5	0.015	33.3
Pb	0.003	0.15	0.02
Sb	<0.0007	0.50	<0.0014
Se	0.001	0.20	0.005
Sr	0.006	3.0	0.002
V	0.07	0.50	0.14
Zn	0.01	4.0	0.003

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

^b Estimates based on average trace element content of fuel oil reported in literature. SSMS analysis performed for the fuel oil samples reports Al and Ca as major components with no numerical values given.

^c By elemental sparging.

TABLE 16. EMISSION FACTORS FOR TRACE ELEMENTS
DURING COAL FIRING

Element	Emission Factor, ng/J		Removal Efficiency, %	Enrichment Factor
	Scrubber Inlet	Scrubber Outlet		
Al	49	1.1	98	1.0
As	0.37	0.35	4	41
Be	0.0079	0.00067	91	3.7
Ca	18	0.73	96	1.7
Cd	1.9	0.21	89	4.8
Co	0.069	0.0047	93	2.9
Cr	0.48	0.046	90	4.1
Cu	0.45	0.072	84	6.9
Fe	150	4.9	97	1.4
Hg	0.035	0.0021	94	2.6
Mn	0.26	0.054	79	8.8
Ni	0.73	0.020	97	1.2
Pb	4.1	1.1	74	11
Sb	0.28	0.099	66	15
Se	0.14	0.033	76	10
Sr	0.17	0.014	92	3.5
V	0.29	0.030	89	4.4
Zn	39	7.7	80	8.4
Total	265	17	94	

TABLE 17. EMISSION FACTORS FOR TRACE ELEMENTS
DURING OIL FIRING

Element	Emission Factor ng/J
Al ^a	0.029
As	<0.0002
Be	<0.0002
Ca ^a	0.567
Cd	<0.0002
Co	0.011
Cr	0.002
Cu	0.002
Fe	0.05
Hg	<0.002
Mn	0.0009
Ni	0.182
Pb	0.0009
Sb	<0.0002
Se	0.0005
Sr	0.002
V	0.023
Zn	0.005

^a Estimated based on average trace element content of fuel oil reported in literature. SSMS analysis performed for the fuel oil samples reports Al and Ca as major components with no numerical values given.

Liquid Waste

There are no significant wastewater streams associated with the oil fired boiler tested. For the coal fired boiler, the two major wastewater streams are: (1) wastewater discharge from the slag tank to the ash pond; and (2) overflow from the settling pond for spent scrubber slurry. The flowrates for these two wastewater streams are approximately 2.89 Gg/hr and 0.77 Gg/hr, respectively.

Inorganics

Analytical results for major inorganic cations in the wastewater stream from the slag tank to the ash pond and the scrubber slurry settling pond overflow are presented in Table 18. Also presented are the analytical results for the scrubber make-up water obtained from the settling pond, and for the scrubber discharge liquid (filtrate from the spent slurry). Of the 18 elements analyzed, iron exceeds its health MATE value and iron, calcium, aluminum, cadmium, vanadium, and zinc exceed their respective ecological MATE values for the wastewater stream to the ash pond. For the scrubber slurry settling pond overflow, calcium, cadmium, manganese, nickel, and lead exceed both their health and ecological MATE values; additionally aluminum, iron, and zinc exceed their respective ecological MATE values. Comparison of the inorganic data for the scrubber slurry pond overflow and the scrubber make-up water (from the settling pond) indicates that the trace element concentrations in these two streams are almost identical. This agreement supports the reliability and accuracy of sampling and analysis of trace elements for the wastewater streams.

Organics

Concentrations of C₇ to C₁₆ organics and high molecular weight (>C₁₆) organics measured in the wastewater streams from coal firing are summarized in Table 19. The total organics detected are low, ranging from 0.06 mg/liter in the wastewater to the ash pond to 0.57 mg/liter in the scrubber slurry settling pond overflow.

GC/MS analyses were performed to identify the organic compounds present in the wastewater streams from coal firing. In the extraction of the aqueous samples, the samples were first acidified to pH 2 and extracted with methylene chloride. The samples were then adjusted to pH 7 and reextracted with methylene chloride. A final extraction with methylene chloride was made at pH 11.

The results of the GC/MS analyses are presented in Table 20. In general, the detected compounds consist of oxygenates such as ketones, alcohols, ethers, and cyclic ethers. Some of these are lightly halogenated. Typical MATE values for these classes of compounds are greater than 1000 µg/liter. Thus, the levels of organics present in the wastewater streams from coal firing do not appear to warrant any environmental concern.

TABLE 18.
TRACE ELEMENT CONCENTRATIONS IN WASTEWATER DISCHARGES FROM COAL FIRING - TEST 135

Element	WASTE Value, mg/l			Water to Ash Pond			Settling Pond Overflow			Scrubber Make-up Water			Scrubber Discharge Liquid		
	Health	Ecology	mg/l	Health	Potential Degree of Hazard	Ecology	mg/l	Health	Potential Degree of Hazard	Ecology	mg/l	Health	Potential Degree of Hazard	Ecology	mg/l
Al	80	1.0	3.5	0.044	3.5	1.5	0.019	1.5	0.015	1.2	0.083	0.0010	0.083		
As	0.250	0.050	0.012	0.048	0.24	0.021	0.084	0.42	0.017	0.066	0.34	0.06	0.24	1.2	
Be	0.030	0.055	0.0003	0.010	0.0055	<0.0008	<0.027	<0.015	0.0091	0.0033	0.0018	0.0011	0.037	0.020	
Ca	240	16	146	0.61	9.1	934	3.9	58	913	3.8	57	384	1.6	24	
Cd	0.050	0.001	0.020	0.40	20	0.059	1.2	59	0.052	1.0	52	0.0005	0.010	0.50	
Co	0.75	0.25	0.002	0.0027	0.0080	0.043	0.057	0.17	0.047	0.063	0.19	0.005	0.0057	0.020	
Cr	0.25	0.25	0.012	0.048	0.048	<0.002	<0.0080	<0.0080	0.001	0.0040	0.0040	0.004	0.016	0.016	
Cu	5.0	0.050	0.008	0.0016	0.16	0.004	0.00080	0.080	0.003	0.00060	0.060	0.004	0.00080	0.080	
Fe	1.5	0.25	3.0	2.0	12	1.1	0.73	4.4	0.63	0.42	2.5	0.0145	0.0097	0.058	
Hg	0.010	0.250	<0.0002	<0.020	<0.00080	<0.00008	<0.0080	<0.00032	<0.00008	<0.00032	<0.00032	<0.00008	<0.00032	<0.00032	
Mg	90	86	13.8	0.15	0.16	N/A	--	--	10.8	1.2	1.2	268	3.0	3.1	
Mn	0.25	0.10	<0.38	<1.5	<3.8	1.9	7.6	19	1.7	6.8	17	<0.38	1.5	3.8	
Ni	0.23	0.010	0.01	0.043	1.0	0.70	3.0	70	0.86	3.7	86	0.20	0.87	20	
Pb	0.25	0.05	0.030	0.12	0.60	0.60	2.4	17	0.024	0.096	0.48	0.03	0.12	0.60	
Sb	7.5	0.20	0.002	0.0003	0.010	0.044	0.0059	0.22	0.041	0.0055	0.21	0.063	0.0084	0.31	
Sr	46	No DATE	0.836	0.018	--	6.4	0.14	--	5.9	0.13	--	3.42	0.074	--	
V	2.5	0.15	0.25	10	1.7	0.064	0.026	0.43	0.080	0.032	0.53	0.095	0.038	0.63	
Zn	25	0.10	0.13	0.052	1.3	2.2	0.088	22	1.7	0.068	17	0.005	0.0002	0.050	

TABLE 19. ORGANICS IN WASTEWATER DISCHARGES FROM COAL-FIRING

	Concentration, mg/l			
	Water to Ash Pond	Settling Pond Overflow	Scrubber Make-up Water	Scrubber Discharge Liquid
C ₇	0	0	0	0
C ₈	0.02	0	0	0
C ₉	0	0	0	0
C ₁₀	0	0	0	0
C ₁₁	0.02	0.04	0.04	0
C ₁₂	0	0.01	0	0
C ₁₃	0	0	0	0
C ₁₄	0	0	0	0
C ₁₅	<0.01	<0.01	<0.01	0
C ₁₆	0.01	0.01	0.01	0
>C ₁₆	0	0.5	0.3	0.1
Total	0.06	0.57	0.36	0.1

TABLE 20. GC/MS ANALYSES OF ORGANICS
IN WASTEWATER STREAMS FROM COAL FIRING

Compound	Water to Ash Pond			Settling Pond		Scrubber Make-up Water		Scrubber Discharge Liquid	
	Acid Extract ^a	Neutral Extract	Basic Extract	Acid Extract	Overflow	Neutral Extract	Basic Extract	Basic Extract	Basic Extract
Olefin or ketone; C ₈ - C ₁₁	X								
Tetrachloropropane (possible)	X								
6-chloro-N-ethyl-N'-(1-methyl ethyl)-1,3,5-triazine-2,4-diamine	X								
8-methyl-3a-dihydronaphthalene-one		X					X		
3a,7a-dihydro-5-methyl-indene-1,7(4h)-dione		X				0.3 µg/l	X		
Quinoline		X							
Butyl naphthalene(4) (plus a possible alkyl substituted naphthalene)					4 µg/l				
1-chloro-2,4-hexadiene					0.5 µg/l				
C ₆ nitrile or C ₅ alcohol							X		
Dl-2-ethyl-hexyl ester of nonane dioic acid						2 µg/l			
2,2,5,5-tetramethyl hexane								X	
Diphenylheptane (possible)									X

^a Identified compounds are present in this extract at concentrations below 15 µg/l.

Solid Waste

There are no significant solid wastes generated from the oil fired boiler tested. For the coal fired boiler, the two major solid waste streams generated are: (1) bottom slag/fly ash from the slag tank; and (2) scrubber sludge from the FGD operation. These two solid wastes are generated at the rates of 110 Mg/hr and 130 Mg/hr on dry basis, respectively.

Inorganics

The concentration of trace elements present in the combined fly ash/bottom slag and in the scrubber sludge are presented in Tables 21 and 22. In about two-thirds of the cases, the trace element concentrations in the combined ash and the scrubber sludge have exceeded the health or the ecological MATE value for solids and in about half of the cases, have exceeded both values. The potential degree of hazard for the trace elements in the combined ash and the scrubber sludge is sufficiently high to warrant the disposal of these solid wastes in specially designed landfills.

Organics

Concentrations of C₇ to C₁₆ organics and high molecular weight (>C₁₆) organics measured in the solid wastes from coal firing are summarized in Table 23. The total organics amount to 86.2 mg/kg for the combined bottom ash/fly ash and 6.6 mg/kg for the scrubber discharge solids. High molecular weight organics were not detected for either solid waste.

Organics present in the bottom ash/fly ash are probably the result of incomplete combustion, or the adsorption of organics by fly ash particulates. Organics are present in the scrubber discharge solids because of the partial removal of these compounds from the flue gas stream in the FDG system. Although no specific organic compound identification information is available, the high trace element content of these solid wastes far outweighs the concern for the organic content. Disposal in specially designed landfills should be satisfactory to handle the potential degree of hazard.

TABLE 21. TRACE ELEMENT CONTENT OF BOTTOM AND
FLY ASH FROM COAL FIRING

Element	Concentration $\mu\text{g/g}$	MATE value, $\mu\text{g/g}$		Potential Degree of Hazard	
		Health	Ecology	Health	Ecology
Al	82,000	16,000	200	5	410
As	10.6	50	10	0.21	1
Be	8.2	6	11	1.40	0.75
Ca	66,000	48,000	3,200	1.40	21
Cd	1.2	10	0.2	0.12	6
Co	44	150	50	0.29	0.88
Cr	208	50	50	4.1	4.1
Cu	822	1,000	10	0.82	82
Fe	174,000	300	50	580	3,480
Hg	< 1	2	50	< 0.5	< 0.02
Mg	ND ^a	18,000	17,400	---	---
Mn	698	50	20	14	35
Ni	328	45	2	7.3	164
Pb	160	50	10	3.2	16
Sb	7.0	1,500	40	0.0046	0.18
Sr	282	9,200	ND	0.03	---
V	86	500	30	0.17	2.87
Zn	1,508	500	20	3.02	75

^a ND - data not available.

TABLE 22. TRACE ELEMENT CONTENT OF SCRUBBER DISCHARGE SOLIDS

Element	Concentration $\mu\text{g/g}$	MATE value, $\mu\text{g/g}$		Potential Degree of Hazard ^a	
		Health	Ecology	Health	Ecology
Al	24,000	16,000	200	1.5	120
As	111.4	50	10	2.23	11.14
Be	2.48	6	11	0.41	0.23
Ca	51,000	48,000	3,200	1.06	16
Cd	36	10	0.2	3.6	180
Co	22	150	50	0.15	0.44
Cr	52	50	50	1.04	1.04
Cu	188	1,000	10	0.19	18.80
Fe	50,000	300	50	167	1,000
Hg	< 1.0	2	50	< 0.5	< 0.02
Mg	0.487	18,000	17,400	< 0.0001	< 0.0001
Mn	564	50	20	11.28	28.2
Ni	96	45	2	2.13	48
Pb	1,080	50	10	21.6	108
Sb	38	1,500	40	0.03	0.95
Sr	994	9,200	ND ^b	0.11	---
V	188	500	30	0.38	6.27
Zn	6,492	500	20	12.98	325

^a Calculation carried to four significant figures.

^b ND - data not available.

TABLE 23. ORGANICS IN SOLID WASTE STREAMS
FROM COAL FIRING

Carbon Number	Concentration, mg/kg	
	Bottom Ash/ Fly Ash	Scrubber Discharge Solids
C ₇	0	0
C ₈	33.0	0
C ₉	28.8	4.7
C ₁₀	0	1.9
C ₁₁	0	0
C ₁₂	0	0
C ₁₃	0	0
C ₁₄	0	0
C ₁₅	0	0
C ₁₆	24.4	0
>C ₁₆	0	0
Total	86.2	6.6

RESULTS

- (1) Average emissions of total particles were 1090 ng/J for the coal fired boiler prior to scrubbing and 7.5 ng/J for the oil fired boiler. Controlled particle emissions from the coal fired boiler tested were 80 ng/J, which were 10 times the particle emissions from the oil fired boiler.
- (2) For the coal fired boiler, over 99 percent of the total particle was greater than 3 μm before scrubbing. The particle size distribution data indicate that approximately 93 percent of the total particles were less than 3 μm after scrubbing.
- (3) For the coal fired boiler, there appeared to be a net increase in emission rates across the scrubber for particles less than 3 μm in size. This net increase can be attributed to the poor removal efficiency of the scrubber for fine particles, the possibility that fine particles may be generated within the scrubber, or that the particle size distribution may be modified during the high energy scrubbing process.
- (4) Stack emissions of inhalable particles ($<15 \mu\text{m}$) were 75 ng/J for the coal fired boiler and 7.5 ng/J for the oil fired boiler.
- (5) Uncontrolled SO_2 emissions from the coal fired and oil fired boilers were 3170 ng/J and 102 ng/J, respectively. Controlled SO_2 emissions from the coal fired boiler were 740 ng/J, which corresponds to a mean scrubber efficiency of 77 percent. Stack emissions of SO_2 from the coal fired boiler were 7 times the SO_2 emissions from the oil fired boiler.
- (6) For the coal fired boiler prior to scrubbing, approximately 1.1 percent of the sulfur compounds in the flue gas were present as SO_3 and 0.4 percent as SO_4 . For the oil fired boiler, approximately 1.2 percent of the sulfur compounds in the flue gas were present as SO_3 and 1.1 percent as SO_4 .
- (7) Stack emissions of SO_3 were 9.5 ng/J for the coal fired boiler and 1.1 ng/J for the oil fired boiler. Stack emissions of SO_4 for the coal fired boiler, at 3.8 ng/J, were also greater than SO_4 emissions of 1.3 ng/J for the oil fired boiler.
- (8) NO_x emissions from the coal fired boiler tested were approximately 6 times the NO_x emissions from the oil fired boiler tested. The differences in NO_x emissions are mainly due to the high heat release rates of the cyclone furnaces for coal firing and the use of NO_x controls for oil firing.
- (9) Total organic emissions from the coal fired boiler were 2.8-4.1 ng/J before scrubbing and 1.4-2.6 ng/J after scrubbing. These were greater than the total organic emissions of 0.4-0.6 ng/J from the oil fired boiler.

- (10) For coal firing, most of the POM compounds found at the scrubber inlet were naphthalene, substituted naphthalenes, biphenyl, and substituted biphenyls. Concentration levels of these POM compounds were several orders of magnitude below their respective MATE values. No POM compounds were identified at the scrubber outlet.
- (11) Total POM emissions from oil firing were lower than those from coal firing before scrubbing. Naphthalene was the principal component of the POM emissions from oil firing. With the exception of the possible presence of benzo(a)pyrene, concentrations of POM emissions from oil firing were also several orders of magnitude below their respective MATE values.
- (12) Of the 18 major trace elements analyzed in the flue gas stream for the coal fired boiler, 16 exceeded their health MATE values at the scrubber inlet and 7 at the scrubber outlet. The seven elements which are of environmental concern from coal firing are arsenic, cadmium, chromium, iron, nickel, lead, and zinc. For oil firing, only chromium and nickel in the flue gas exceed their health MATE values.
- (13) Two major wastewater streams from coal firing were not associated with oil firing. These were the wastewater discharge to the ash pond and the overflow from the scrubber slurry settling pond. Analysis of these wastewater streams indicates that the concentration levels of some of the trace elements present exceeded their respective health and ecological MATE values. However, the concentration levels of organics present in these wastewater streams do not indicate any major environmental concerns currently.
- (14) No significant solid wastes were generated from the oil fired boiler. For the coal fired boiler, the two major solid waste streams were the combined bottom slag/fly ash from the slag tank and the scrubber sludge. For these two solid wastes, almost all the 18 trace elements analyzed have exceeded both their health and ecological MATE values. Because the trace elements may leach from the disposed ash and scrubber sludge, these solid wastes must be disposed of in specially designed landfills.
- (15) In summary, the controlled multimedia emissions from the coal fired boiler tested were of greater environmental concern than the multimedia emissions from the oil fired boiler tested. Stack emissions of pollutants from coal firing could be further reduced by proper NO_x control and higher pressure drop across the venturi scrubber for more efficient SO₂ and particulate removal. However, the reduction in gaseous emissions from coal firing could also result in higher pollutant levels in the solid wastes generated at the coal fired power plant. The consequence will be the transferring of a high volume, low concentration pollution stream to a low volume, high concentration stream which can be more readily contained.

ACKNOWLEDGEMENT

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TITLE

OPERATING AND STATUS REPORT
WELLMAN-LORD SO₂ REMOVAL/ALLIED CHEMICAL SO₂ REDUCTION
FLUE² GAS DESULFURIZATION SYSTEMS²
AT
NORTHERN INDIANA PUBLIC SERVICE COMPANY
AND
PUBLIC SERVICE COMPANY OF NEW MEXICO

by

D. W. Ross
Director of Technical Services
Davy Powergas Inc.
Lakeland, Florida

James Petrie
SO₂ Chemical Plant Supervisor
Public² Service Company of New Mexico
San Juan Station
Waterflow, New Mexico

F. W. Link
Supervisor Plant Engineering
Northern Indiana Public Service Company
Michigan City, Indiana

A paper presented at the Fifth Symposium on Flue Gas Desulfurization
for the Environmental Protection Agency

March 5-8, 1979
Las Vegas, Nevada

ABSTRACT

This paper has been written in two Sections.

Section I contains an up-to-date report on the FGD plant at NIPSCO's Dean H. Mitchell Station at Gary, Indiana, and the most recent data pertaining to plant operations, problems encountered with corrective actions and anticipated plans for future plant operations. The FGD plant consists of the Davy/Wellman-Lord SO_2 Recovery Process and Allied Chemical Corporation's SO_2 Reduction Process. The plant information starts after the successful completion of the performance test on September 16, 1977, and continues to January 1, 1979.

Section II has a report on Public Service Company of New Mexico, at Waterflow, New Mexico. This plant also consists of the Davy/Wellman-Lord SO_2 Recovery Process and Allied Chemical Corporation's SO_2 Reduction Process. The Wellman-Lord SO_2 Recovery Process is attached to power generation Units No. 1 and 2, and is composed of three operating sections: (1) Separate SO_2 Absorption for each power generation unit, (2) a dual train Chemical Regeneration Plant and single Purge Treatment Unit, and (3) a dual train Allied Chemical SO_2 Reduction Unit. This part of the report contains information beginning with the initial plant start-up in April 1978 and follows to January 1, 1979. It too will outline the plant operations, problems encountered with corrective action taken, and plans for continued plant operations.

INTRODUCTION⁽¹⁾

The Wellman-Lord SO₂ Recovery Process was developed in the late 1960's. The process is being used throughout the world, Exhibits 1 and 2, pages 53 and 54. At every installation, the Process has proven itself as operationally reliable and has met or surpassed all governmental regulations regarding SO₂ emissions.

The Process may be applied to any flue gas containing SO₂. It may be applied to the flue gas from all fossil fuel fired boilers, nonferrous smelters, sulfuric acid plants, and Claus plants. The Process uses a Wellman-Lord recycle system and yields an SO₂ gas suitable for conversion to elemental sulfur as demonstrated in the Northern Indiana Public Service Company's D. H. Mitchell Plant and Public Service Company of New Mexico's San Juan Station using Allied Chemical technology for the reduction process.

PROCESS DESCRIPTION

Wellman-Lord SO₂ Recovery⁽²⁾

The Wellman-Lord process consists of three major operating sections - SO₂ absorption, purge treatment and SO₂ regeneration.

In the SO₂ absorption section, residual fly ash in the flue gas is removed by water scrubbing. SO₂ is then removed from the flue gas by scrubbing with a solution of sodium sulfite. The chemicals contained in this solution remain completely dissolved throughout the absorber. Flue gas scrubbing with a clear solution, free from suspended solids, plugging and scaling, is a fundamental reason underlying the exceptional on-stream reliability experienced in the commercial operations of the Wellman-Lord process. A general schematic flow diagram of the process is shown in Figure 1, page 4.

The purge treatment section selectively removes inactive oxidized sodium compounds from a sidestream of the absorbing solution and converts this material into a dry granular product which is marketed.

The third section of the Wellman-Lord process involves thermal regeneration of the absorbing solution to release the absorbed SO₂ as a concentrated gas stream and return of the reconstituted solution to the absorber.

The concentrated SO₂ gas may be converted to liquid SO₂, sulfuric acid or elemental sulfur. NIPSCO and PNM elected to use the Allied Chemical's SO₂ Reduction Process. A general schematic flow diagram of the process is shown in Figure 1, page 4.

Allied Chemical SO₂ Reduction to Sulfur

Sulfur is recovered by Allied Chemical's SO₂ reduction process which consists of two principal operating sections.

In the primary reduction section, more than one-half of the entering SO₂ is converted to elemental sulfur. A key feature of this section is the effective control of chemical reactions between SO₂ and natural gas over a catalyst developed by Allied Chemical for this purpose. Heat generated by these chemical reactions is recovered and utilized to preheat the SO₂ gas stream entering this section.

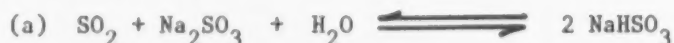
Packed bed regenerative heaters provide a rugged and efficient means for achieving this heat exchanger function. The process gas flow through the regenerators is periodically reversed to alternately store and remove heat from the packing; hence, the overall section is thermally self-sustaining.

Automatic control of the flow reversing cycles and other process conditions achieves optimum performance in the system, with high sulfur recovery efficiency and reductant utilization at all operating rates.

The gas leaving the primary reactor system is cooled in a sulfur condenser, for condensation and recovery of sulfur product. The remaining gas, containing proper proportions of SO_2 and H_2S is processed through a Claus conversion system for recovery of additional sulfur product. The Claus system off-gas is incinerated and recycled to the Wellman-Lord SO_2 absorber.

PROCESS CHEMISTRY

The Wellman-Lord Process is based on the chemistry of the sodium sulfite/bisulfite system: flue gas containing SO_2 is scrubbed with a sodium sulfite solution which absorbs SO_2 , converting sodium sulfite to sodium bisulfite:

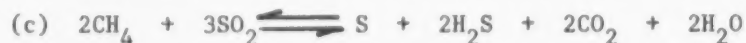


The sodium bisulfite solution is regenerated by thermal decomposition. Application of heat simply reverses equation (a):



The SO_2 is recovered in a concentrated stream.

The concentrated stream of SO_2 gas is then reduced to high purity elemental sulfur in the Allied Chemical Process. This conversion is carried out in two steps. In the first step, a portion of the SO_2 in the feed gas reacts with natural gas, yielding a mixture of elemental sulfur, hydrogen sulfide, carbon dioxide and water vapor:



In the second step, H_2S formed in the first step reacts with the remaining SO_2 yielding additional elemental sulfur and water vapor:



The tail gas from the sulfur plant is incinerated and recycled to the Wellman-Lord absorber.

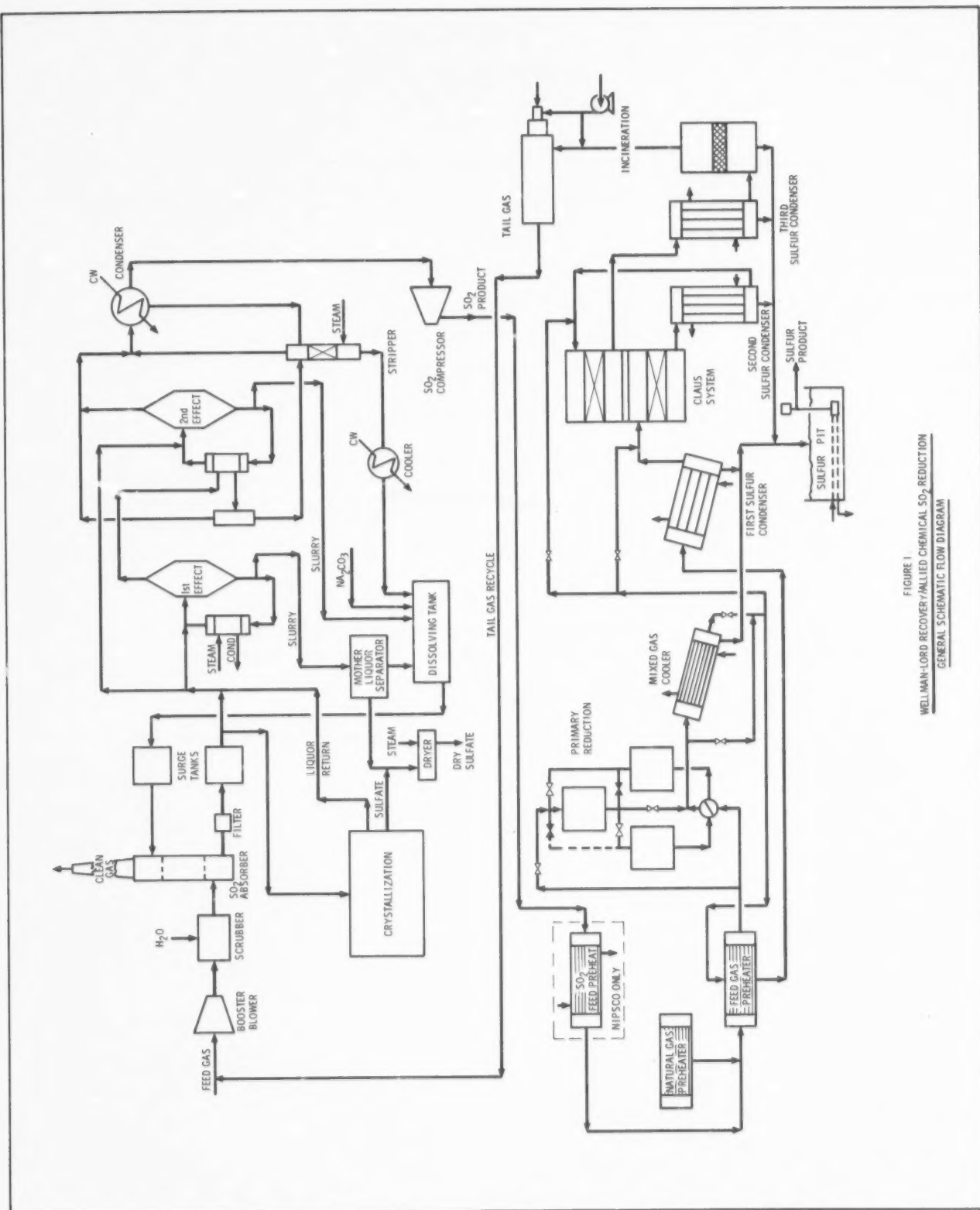


FIGURE 1.
WELLMAN-LORD RECOVERY/ALKALI CHEMICAL SO_2 REDUCTION
GENERAL SCHEMATIC FLOW DIAGRAM

NIPSCO

SECTION I

The first part of this paper, Section I, deals with the Unit No. 11 Boiler FGD System Northern Indiana Public Service Company's Dean H. Mitchell Station located in Gary, Indiana.

This FGD system will be referred to as the NIPSCO FGD Plant in this paper. By definition, the NIPSCO FGD Plant includes the flue gas booster fan, flue gas isolation damper and the flue gas louver bypass damper, all of which are outside battery limits as well as the primary battery limits portion of the FGD plant consisting of the pre-scrubber, the Wellman-Lord absorption, regeneration and purge treatment units and the Allied Chemical SO_2 Reduction Unit. Byproduct storage and loading facilities are included within battery limits.



PLANT Flue Gas Scrubbing System 118 MW Coal-Fired Unit
CLIENT Northern Indiana Public Service Co. &
U. S. Environmental Protection Agency
PROCESS Wetman-Lord/Alised Chemical
PRODUCT Elemental Sulfur
CONTRACTOR Davy Powergas Inc. Lakeland, Florida

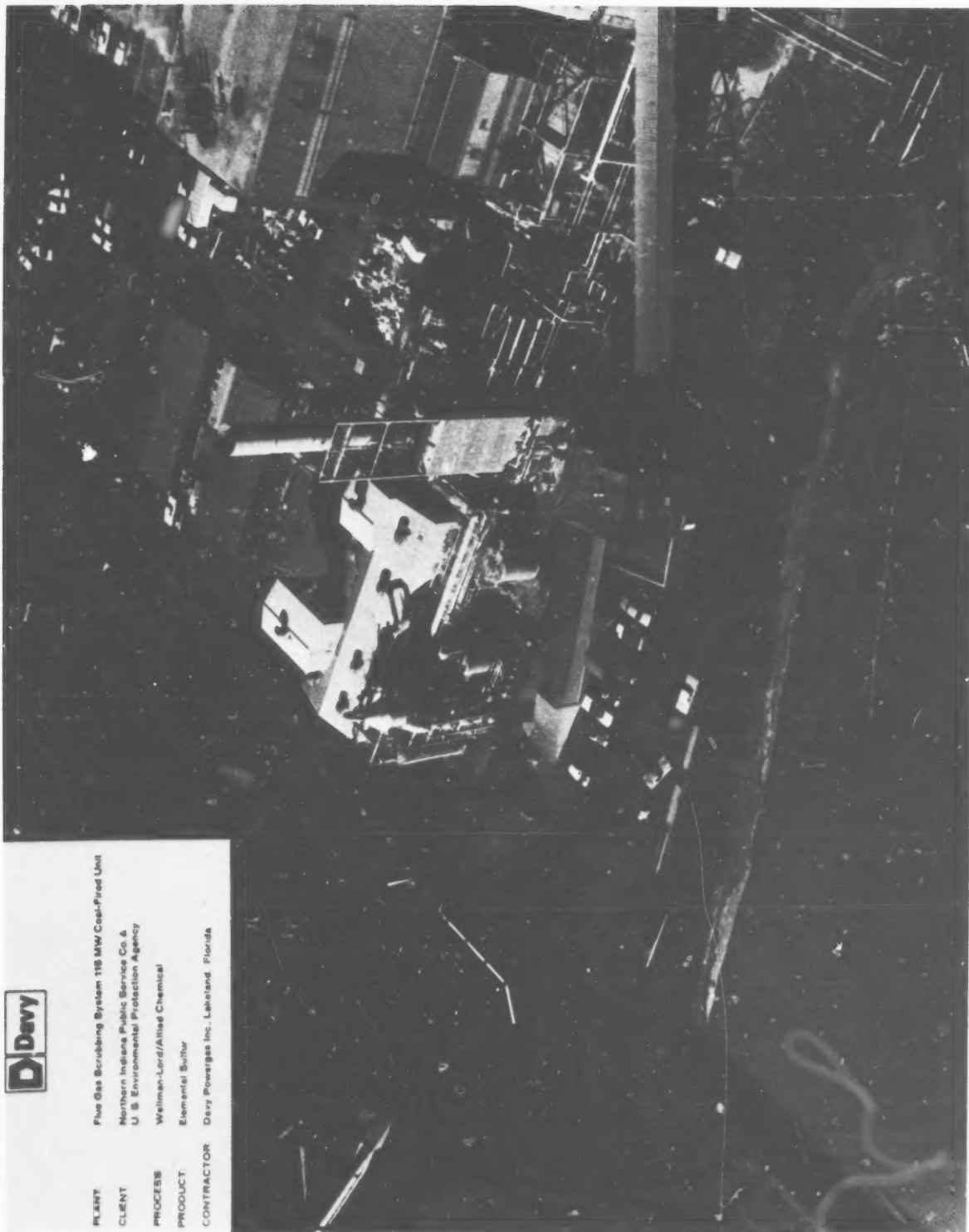


Figure 2
GENERAL VIEW OF NIPSCO PLANT

1. NIPSCO Project Background

NIPSCO and the U.S. Environmental Protection Agency entered into a cost shared contract in June of 1972 for the design, construction and operation of a regenerable flue gas desulfurization (FGD) demonstration plant. The system selected for the project was a combination of the Wellman-Lord SO₂ Recovery Process and the Allied Chemical SO₂ Reduction Process. The FGD plant was to be retrofitted to NIPSCO's 115 MW pulverized coal-fired Unit No. 11 at the Dean H. Mitchell Station in Gary, Indiana. NIPSCO entered into contracts with Davy Powergas Inc. for the design and construction of the FGD plant and with Allied Chemical Corporation for operation of the plant.

A successful performance test was run from August 29, 1977, through September 16, 1977.

A one-year demonstration test period began on September 16, 1977, and has now been extended to March 15, 1979. A decision is to be made prior to March 15, 1979, regarding further extension of the demonstration test period. (4)

TRW under contract to the U.S. Environmental Protection Agency, has continued to monitor and report the performance of the boiler and the FGD plant during the demonstration test period.

This section of the paper will discuss the FGD plant operating experience and the operating and maintenance costs from the beginning of the demonstration test period on September 16, 1977 through December 31, 1978.

NIPSCO is continuing to assess the Wellman-Lord/Allied Chemical option for SO₂ emission control.

NIPSCO is still optimistic that longer periods of continuous operation will be achieved during the extended demonstration test period. Through continued operations at NIPSCO it will be possible to make a more accurate evaluation of the economics and plant reliability.

2. NIPSCO FGD Performance Design Criteria & Results (1) (3)

During the plant acceptance test, from August 29, 1977, through September 14, 1977, the FGD system performance criteria and obtained results were:

92 MW Equivalent Test

SO₂ Removal

1. Required: Minimum SO₂ removal of 90%, measured continuously and averaged every 2 hours for a period of 83 hours.

2. Results: SO₂ removal averaged 91% over the 12-day test period. In only two 2-hour periods (out of 144) was the SO₂ removal less than 90%, and for those periods, it averaged 88% and 89%.

Particulate Removal

1. Required: Particulate emission measured once daily will not exceed the Federal NSPS for fossil fuel fired steam generators of 0.1 lb/million Btu heat input.
2. Results: Particulate emission averaged 0.04 lb/million Btu, or 40% of the maximum allowable. Of the 12 days, tests could not be run on four days due to inclement weather. On one day, the test data was not valid.

Soda Ash Consumption

1. Required: Average over the 12-day test period not to exceed 6.6 STPD.
2. Results: Soda ash consumption determined by daily inventory obtained from storage bin measurement (official result) averaged 6.2 STPD, or 94% of the maximum allowable. Consumption determined by manual weighing the feeder output every two hours throughout the 12-day test period averaged 5.7 STPD, or 86% of the maximum allowable.

Aggregate Cost of Steam, Electricity and Natural Gas

1. Required: Aggregate cost not to exceed \$56 per hour based on predetermined unit cost.
2. Results: Hourly cost averaged \$43 per hour over the 12-day test Period, or 77% of the maximum allowable.

Sulfur Purity

1. Required: Minimum sulfur purity 99.5%, suitable for conversion to quality sulfuric acid by standard production practice.
2. Results: Sulfur purity determined from a composite sample collected over the 12-day test period was 99.9%, easily exceeding the required purity.

110 MW Equivalent Test

SO₂ Removal

1. Required: Minimum SO₂ removal of 90%, measured continuously and averaged every 2 hours for a period of 83 hours.
2. Results: SO₂ removal averaged 91% over the 3-1/2 day test period. In only one 2-hour period (out of 42) was the SO₂ removal less than 90%, and for that period, it averaged 89%.

Particulate Removal

1. Required: Particulate emission measured once daily will not exceed the Federal NSPS for fossil fuel fired steam generators of 0.1 lb/million Btu heat input.
2. Results: Particulate emission averaged 0.04 lb/million Btu, or 40% of the maximum allowable. Of the 3-1/2 days, a test could not be run on one day due to inclement weather.

Viability of the NIPSCO FGD System (1) (3)

During the 12-day performance period at the 92 MW equivalent rate, there was a total of 26 hours in interruptions in the fully integrated operation of the FGD system. Of the 26 hours, 18 were related to boiler problems and 8 were related to problems in the FGD system. In addition, there was a 4-hour period in which the SO₂ removal averaged "only" 88.5%; this 4-hour period was added to the performance test at the end of the 12-day test period. It should be mentioned that outages in the FGD system did not interrupt SO₂ removal. Furthermore, an SO₂ removal of 88.5% at the NIPSCO site results in an emission well below the NSPS of 1.2 lb/million Btu heat input. During the acceptance test the parameters used by EPA for judging the viability of a FGD system were: availability, 94%; reliability, 100%; operability, 100%; utilization factor, 94%.

3. Operating Experience

The monthly operating hours as shown on charts 1"a" through "p" on pages 15 through 30 illustrate the operating periods of NIPSCO's Unit No. 11 and of the FGD plant. In connection with the bar graphs the following definitions were used:

- a) A solid line indicates that Unit No. 11 or the FGD plant was in operation.
- b) The definition of Unit No. 11 being in operation is: Unit synchronized on line regardless of the megawatt load.

- c) The definition of the FGD plant being in operation is:
- Receiving flue gas from Unit No. 11.
 - No SO_2 is being returned from the evaporator to the existing Unit No. 11 stack.
- d) The Unit No. 11 operating conditions required to permit FGD plant operation are:
- Unit No. 11 operating on high sulfur coal at 46 GMWE minimum load.
 - Sufficient main steam available (530 psig minimum).
 - Sufficient demineralized make-up water available.
 - Unit No. 11 supplied utilities available (electricity, boiler feed water).

4. Problems Causing FGD Outages

Chart 2 on page 31 illustrates the FGD Plant Operating Factor based on hours of FGD operation divided by hours of generating unit operation. Table 1 on page 32 through 34 summarizes the FGD outages as shown on the monthly bar graphs and indicates whether the outage was attributable to the boiler, the FGD plant or a combination of both. At the bottom of Table 1 are listed four categories of major FGD outages. These outages are discussed here.

5. Booster Fan

Problem:

Imbalance of the air foil flue gas booster fan has been a continuous problem since early in the demonstration period. Operating conditions at or below the acid dew point of the flue gas (and below the specified design temperature of 300°F) plus an ineffective guillotine isolation damper upstream of the booster fan contributed to an accumulation of fly ash, water and ice in the fan housing and on the fan blades. Over a period of several months during which the FGD system (and the booster fan) was idle for substantial periods because of high silica problems in Unit No. 11 make-up water, these conditions resulted in corrosion and erosion of the air foil blades which finally required a complete reblading of the fan. Turbine governor malfunctions have apparently been caused by exposure of the governor to outdoor weather and dust conditions.

Solution:

The following revisions and additions are in progress or have been made to resolve the booster fan problems:

- a) The intermediate layer of Unit No. 11 air preheater elements was removed to raise the flue gas exit temperatures.

- b) The booster fan inlet and outlet duct and the booster fan housing were insulated.
- c) The booster fan was rebladed and Inconel shields were installed on the blade leading edges.
- d) A steam soot blower was installed in the booster fan for on line blade cleaning.
- e) A booster fan drive turbine enclosure is presently under construction.

Results:

Preliminary results are very encouraging. The air preheater element removal has resulted in an increase of approximately 30°F for the exit flue gas temperature with no apparent adverse effects on the boiler other than a slight decrease in boiler efficiency. The flue gas temperature is now consistently at or above 300°F and condensation has not occurred in the ductwork and booster fan as it did in the past. A complete assessment of the results of the above on long term fan performance will be made after the plant has operated throughout this winter.

6. High Silica Levels in Boiler

Problem:

In October 1977, routine Unit No. 11 boiler water chemistry tests indicated high levels of silica. The first attempts to solve the problem focused on inspection of Unit No. 11 boiler and condenser for leaks. Frequent or continuous blowdown of the boiler and reduced steam pressures made the steam supply to the FGD plant so unreliable that it precluded integrated FGD operation. After extensive investigation including ultrasonic condenser tube testing, the condenser was found not to be the cause of the contamination. The causes of the silica build-up appear to have been a boiler chemistry upset due to high condensate make-up requirements imposed on the boiler by the FGD plant requirements. This was compounded by the continuous boiler blowdown and lack of continuous silica monitoring on a portable demineralizer being used to supplement the station demineralizer to supply Unit No. 11 and FGD condensate make-up demand.

The need for using a portable unit resulted from failure of a reverse osmosis unit, installed by NIPSCO, to operate successfully.

The design of the FGD system was based on the use of filtered lake water for use as flushing water to the packing gland on the process pumps. During plant start up it was found that the quantity of fine silt in the lake water made it impossible to adequately filter the water for packing gland flushing. As an expedient the use of condensate as a flushing fluid was implemented and the resulting consumption of condensate in the FGD system became substantially greater than initial design.

Since the water treating facilities (station demineralizer and reverse osmosis unit) available at the NIPSCO plant were expected to have adequate capacity, this increase in condensate consumption was not seen as a problem at the time except from the cost standpoint. The unit cost for treating water with the portable demineralizer equipment is very high.

Solution:

A continuous silica analyzer was installed on the portable demineralizer outlet and the silica content of the treated make-up water was limited to 10 parts per billion. Condensate from the FGD system is diverted to waste at a low pH reading of 6.5 or a high reading of 9.5 and at a conductivity reading of 5.5 micromohs/cm and higher. The FGD plant condensate return diverter system was changed from manual to an automatic reset mode. If the condensate is diverted to waste because of high conductivity, low or high pH, it will automatically return to the Unit No. 11 condenser when the pH and conductivity readings are again within acceptable limits. As part of the solution to this problem, a comprehensive program was undertaken in the FGD plant to reduce condensate consumption and losses.

Results:

The above modifications have been effective in controlling Unit No. 11 boiler silica. Combined Unit No. 11 and FGD condensate make-up is now generally in the 40 to 50 gpm range with an occasional excursion to higher volumes if the return condensate has been diverted to waste.

7. Guillotine Isolation Damper

Problem:

A top entry guillotine damper was installed between the outlet of the existing Unit No. 11 boiler induced draft fans and the FGD plant booster fan inlet. During operation with the damper open a build-up of fly ash occurred in the bottom and lower side channels of the damper frame. When the damper was closed it would strike the fly ash build-up and the bottom seals of the damper sustained mechanical damage. Considerable corrosion damage has also been experienced with the damper.

Solution:

Considerable work was done on maintenance and modification of the damper. A bottom channel purge air system was installed and modified. Damper seals were replaced with seals made of a more corrosion resistant alloy.

Results:

The fly ash build-up continued to occur at the bottom of the damper. A decision was made not to invest any additional money in the damper. Manual slide gate dampers were installed in each of the booster fan inlet housings and will be used to isolate the fan when necessary for maintenance.

Note: During the recent period, October through December 1978, the guillotine isolation damper has operated successfully but has not provided total shut-off capability, probably caused by fly ash damaged seals. The recently completed modifications to the Unit No. 11 air preheaters and installation of thermal insulation on the flue gas duct work have allowed the flue gas temperatures to be maintained above the dew point and alleviated some of the problems with the guillotine isolation damper as well as the flue gas booster fan.

8. Coal Quality

Problem:

Wet, poor quality coal caused erratic Unit No. 11 boiler operation. As a result, steam pressure and flow to the FGD plant fluctuated and prevented sustained integrated FGD operation.

Solution:

NIPSCO's coal procurement department negotiated with suppliers to insure a consistent supply of a more acceptable quality coal.

Results:

The coal quality has improved and is no longer considered a problem.

9. Problems Not Causing FGD Outages

There were some problems experienced during the early plant operations which are mentioned here. These situations did not cause FGD Plant outages.

Absorber Tray Leakage

Sodium balances and analyses of the fly ash sump discharge indicated that there were solution losses occurring in the absorber. Inspection of the absorber lower collector tray showed that solution was splashing over the chimney free-boards and leakage was also occurring between the tray perimeter and the absorber wall. The chimney free-board height has now been raised and the tray perimeter recaulked and sealed with rubber.

Evaporator Circulating Pump Driver

The evaporator slurry circulating pump was originally equipped with a steam turbine drive, utilizing main steam supplied by Unit No. 11 boiler to the FDG plant, at 550 psig. Exhaust steam from the drive turbine was utilized as part of the heating steam requirements for the evaporator.

During the demonstration period the main steam supply to the FGD plant was partially or totally interrupted on many occasions resulting in slow down or stopping of the circulating pump. To avoid settling of solids and plugging of the evaporator it was necessary to dilute the solution and dump the evaporator slurry into a holding tank. After the steam supply was re-established, two days would be required to refill and re-establish solids content before evaporator operation could be resumed.

To avoid these lengthy interruptions to the FGD operation an electric motor drive was installed on the pump in September 1978.

Purge Dryer Capacity

System sulfate inventories and dry sodium salt production have recently indicated that the purge dryer possibly was not operating at rated capacity. A test program has been undertaken by Davy Powergas. Mechanical modifications are in progress aimed at increasing the purge dryer capacity.

Chart 1 (a)

NIPSCO FGD PLANT
OPERATING HOURS

SEPTEMBER 1977				
DATE	FGD PLANT		UNIT NO. 11	DATE
1		Acceptance Test in Progress.		1
2				2
3				3
4				4
5				5
6				6
7				7
8				8
9				9
10				10
11				11
12				12
13				13
14				14
15	360 HRS.	Variable Main steam Pressure From Unit 11.		15
16				16
17				17
18	67 HRS.	Below 46 GMW Operation On Unit 11 Due to Wet Coal.		18
19				19
20				20
21				21
22			OFF 7:42P Wet Coal - ON 8:13P Mills Plugged.	22
23			OFF 9:44P Control Valve Problems. ON 10:29P	23
24			OFF 12:50P ON 3:19P	24
25			OFF 11:11P ON 6:43P	25
26			OFF 6:54P ON 7:03P	26
27				27
28			OFF 7:37P Blr. Feed Pump Trip Due ON 9:30P to water From SO ₂	28
29			Feedwtr Rel. Vlv.	29
30				30

Chart 1 (b)

NIPSCO FGD PLANT OPERATING HOURS

OCTOBER 1977			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Kept Shutdown Through Oct.		1
2	5 To Run Heat Balance & Flue Gas Flow Rate Tests		2
3	on Unit 11		3
4			4
5	Start Up of Absorber /Evap. System		5
6			6
7	Instrument Problems in SO ₂ Reduction System.		7
8	Incinerator Repairs, Tube Leaks in No. 1 Sulfur Condenser.		8
9			9
10			10
11			11
12			12
13			13
14			14
15			15
16			16
17			17
18	132 HRS.		18
19	Bearing Failure on steam Turbine Drive		19
20	for Evaporator circul- ating pump		20
21			21
22	Evaporator Started and Held In Stand-By Pending		22
23	Unit 11 Boiler Tube Repair		23
24			24
25			25
26			26
27			27
28	Absorber and Evaporator Operating.		28
29		8:24 PM Boiler Tube Leak & Check for Condenser	29
30		Leaks.	30
31		7:46 AM	31

Chart 1 (c)

NIPSCO FGD PLANT
OPERATING HOURS

NOVEMBER 1977				
DATE	FGD PLANT		UNIT NO. 11	DATE
1	8 HRS.	■	■	1
2		Booster Fan Cleaning and Balancing.		2
3	22 HRS.	■		3
4		↓		4
5		■		5
6		■		6
7		■		7
8		■		8
9		■		9
10		■		10
11		■		11
12		■		12
13		■		13
14		■		14
15		■		15
16		■		16
17		■		17
18		■		18
19		■		19
20		■		20
21		■		21
22	428 HRS.	■		22
23			8:14 PM ■ Inspect Steam Side of Condenser	23
24				24
25				25
26		Guillotine Damper Jammed In Part	7:54 AM ■	26
27		Open Position.		27
28		Gasket in Solution		28
29		Line From Evaporator to Circ. Pump Blew Out.		29
30		↓	■	30

Chart 1 (d)

NIPSCO FGD PLANT OPERATING HOURS

DECEMBER 1977				
DATE	FGD PLANT	UNIT NO. 11		DATE
1	FGD Plant Held In Standby Pending			1
2	Resolution of High Silica Problem on			2
3	Unit 11 Boiler.			3
4				4
5				5
6	NOTE: After extensive investigation, it was determined the high			6
7	Silica Problem on the			7
8	Unit 11 Boiler was caused by high			8
9	condensate make-up requirements imposed on	12:58 AM	SO ₂ Guillotine Damper	9
10	Boiler by the FGD System. Normal Boiler make-up without FGD is		Repair and Condenser	10
11	approx 15 GPM. Boiler make-up with FGD in		Inspection and clinker	11
12	operation is normally	1:10 AM	Repair Air Circuit	12
13	approx 40 to 50 GPM but can increase to 150 GPM	10:51 PM	Breaker on #11	13
14	during FGD start up or upset operating modes.	3:22 AM	Disconnects.	14
15				15
16	High silica in the make-up water to Unit 11	10:50 PM	Repair #3 Precip.	16
17	Boiler was corrected and subsequently maintained under control		Field and Check	17
18	by NIPSCO through:	12:48 AM	Condenser for Leaks.	18
19	a. Installation of a	7:58 AM	Loss of Control Air	19
20	continuous silica analyzer on the	8:35 AM		20
21	treated make-up water to permit			21
22	adequate control of the demineralizer			22
23	operation.		S.D. For Coal Mill repair	23
24	b. Installation of additional controls			24
25	on the return condensate diverting equipment			25
26	to avoid unnecessary and prolonged			26
27	diversion of condensate to waste.			27
28				28
29	c. A comprehensive program was undertaken to reduce the			29
30	condensate consumption and losses.			30
31				31

Chart 1 (e)

NIPSCO FGD PLANT
OPERATING HOURS

JANUARY 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	FGD Plant held in Standby Pending Res-		1
2	olution of High Silica		2
3	Problem on Unit 11		3
4	Boiler		4
5			5
6		8:45 PM Repair Precip. Fields	6
7		and check condenser	7
8		for leaks.	8
9		8:40 AM	9
10			10
11			11
12			12
13			13
14		Condenser Tested for	14
15		Leaks	15
16			16
17	FGD Plant isolated		17
18	from Unit 11 Steam		18
19	Supply, Boiler Feed-		19
20	water Supply, and		20
21	Condensate Return.		21
22		9:30 PM Repair Precip. Field	22
23		and 11-3 Mill Hot	23
24		Air Damper	24
25		3:31 AM	25
26	FGD Plant Steam		26
27	Supply Changed back		27
28	to Unit 11 Main		28
29	Steam. SO ₂ comp-		29
30	ressor gasket leak.		30
31			31

Chart 1 (f)

NIPSCO FGD PLANT
OPERATING HOURS

FEBRUARY 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Booster Fan Cleaning Complete Additional Access Doors and Drains Installed in Fan Housing.		1
2			2
3	FGD Start-UP Held Pending Unit #11	3:42 AM	3
4	Boiler Tube Repair	8:45 AM 9:14 PM	4
5	Clean Duct Downstream from Guillotine Damper.	Trip on High Furnace Pressure Repair Boiler Tube Leak	5
6	Damper Push Rods broke when attempt was made to open Damper.	2:50 AM	6
7			7
8			8
9		10:40 PM	9
10	Repair started on Guillotine Damper.	Stop Valve Repair	10
11			11
12			12
13		Stop Valve Work Complete Unit kept down to	13
14		Finish SO ₂ Damper Repair	14
15			15
16	Damper Repair Complete at 2:28 A.M.	10:14 AM 11:28 PM	16
17	Booster Fan started on slow Roll. Shaft was bowed from Standing.	2:47 AM 3:08 AM	17
18	Air Line to Governor frozen. Fan rebalanced. Flue Gas admitted to Absorber. Booster Fan Oil Cooler frozen - lost Bearing Oil. Refilling Oil System.	3:24 AM	18
19	Booster Fan and Absorber in Operation.	Turbine Pre-emergency Governor Test caused Unit to Trip.	19
20			20
21	Evaporator Circulating Pump down for Repacking.	Limited to 55 GMW due to I.D. Fan Bearing Problem.	21
22	Frozen SO ₂ Superheater Line.		22
23		Switched to Low Sulfur Coal Due to Feeding Problems with High Sulfur Coal.	23
24			24
25	Evaporator Circulating Pump Packing Problem.		25
26			26
27			27
28			28

Chart 1 (g)

NIPSCO FGD PLANT
OPERATING HOURS

MARCH 1978				
DATE	FGD PLANT		UNIT NO. 11	
1		Booster Fan Vibration - Fan Re-Balanced.		Coal Problems - Low BTU Content Wet.
2		Booster Fan and Absorber Started Up. Main steam pressure to FGD fluctuating down to 250PSIG		Full of Rocks.
3		Waiting for SO ₂ Levels to Build-UP so Reduction Section can be started.		
4				Coal Feed Problems Load below 46 GMWE
5				Precipitator Reduced Efficiency.
6	1:15 P.M.	Started Sulfur Production.		Two coal mills out of Service 50 GMW Load
7				
8		Main Steam Pressure to FGD fluctuating down to 300 PSIG		Coal Mill problem in AM
9				
10				
11				
12				
13				
14				
15	12:00 N	Unit shutdown due to low main steam pressure.	10:34 PM	Load below 46 GMWE - One coal mill in operation
16		Guillotine isolation damper did not close completely		Precipitator wire repair.
17		Balance Booster Fan Start Up	10:04 PM	Repair ash hopper.
18		Booster Fan and Absorber		Trouble Shoot Turbine Pre-emergency governor
19		Guillotine isolation damper jammed. Opened after manual assist.		
20				
21				
22				
23		Booster Fan Balanced Again. Absorber Started Up.	2:35 PM 6:05 PM	Guillotine Damper would not close.
24				Unit shutdown to balance booster fan.
25				
26		Instrument Problems		
27				On Low Sulfur Coal due to Feeding problems with Wet High Sulfur Coal.
28		Booster Fan Vibration Fan on Slow roll		I.D. Fan Vibration Problem. Fan Shutdown
29		pending arrival of service engineer.		pending Unit 11 outage.
30				
31		Clean Guillotine damper bottom channel.	8:51 PM	Repair & Balance West I.D. Fan.

Chart 1 (h)

NIPSCO FGD PLANT
OPERATING HOURS

APRIL 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1		Repair & balance West I.D. Fan Inspect	1
2		guillotine isolation damper and duct.	2
3	Service Engineer to advise Re: Booster fan	4:13 AM	3
4	balancing.		4
5	Booster fan blades re-		5
6	placed due to corrosion and erosion of existing blading.		6
7			7
8			8
9			9
10			10
11			11
12			12
13		10:25 PM	13
14		Repair I.D. fan blading and repair No. 3 precipitator field.	14
15			15
16			16
17			17
18			18
19		1:30 AM	19
20			20
21			21
22			22
23			23
24		10:55 AM	24
25		11:45 AM	25
26	Absorber lower collector tray was recaulked and loose rubber was removed.		26
27	Booster fan reblading complete Fan ready for balancing.		27
28	Balanced booster fan.		28
29	Guillotine isolation damper inoperable.		29
30			30

Chart 1 (1)

NIPSCO FGD PLANT
OPERATING HOURS

MAY 1978				
DATE	FGD PLANT		UNIT NO. 11	DATE
1		Plant on stand-by awaiting opening and		1
2		repair of guillotine		2
3		isolation damper		3
4		Guillotine isolation damper blocked in full open position	8:37 PM	4
5			Shutdown for repair of SO ₂ plant guillotine isolation damper. Two coal mills out of service.	5
6		Booster fan and absorber started up.	10:35 AM	6
7		Evaporation started but SO ₂ flow rock too low		7
8		for operation of SO ₂ Reduction Unit.		8
9				9
10				10
11	7:30 PM	Sulfur Production started.		11
12				12
13		Boiler feedwater booster pump failure. No BFW supply available for the FGD System.	Changed to low sulfur coal because of wet H.S. coal.	13
14	1:20 PM	SO ₂ levels too low for operation of reduction sections	Resumed high sulfur coal feed to bunkers.	14
15				15
16				16
17				17
18		Alternate boiler feedwater tie-in complete.		18
19	12:00 N	BFW supply now at temperature substantially below required 228°F.		19
20				20
21				21
22				22
23				23
24				24
25				25
26				26
27	11:00 PM	Reduction area shutdown due to high pressure drop through coalescer.		27
28				28
29				29
30				30
31				31

Chart 1 (J)

NIPSCO FGD PLANT
OPERATING HOURS

JUNE 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Continuing to repair and clean coalescer completed June 2.		1
2	Main steam to FGD shut down due to pressure reducing valve problem.		2
3	Erratic pressure on main steam supply to FGD plant.		3
4			4
5			5
6			6
7	Booster fan turbine bearing failure.		7
8	Booster fan turbine repair in progress.		8
9			9
10			10
11			11
12			12
13			13
14			14
15			15
16			16
17			17
18			18
19	Started removal of catalyst in A&B claus converters.		19
20	Replacement complete		20
21	on 6-23 in PM.		21
22	Booster fan turbine repair complete. Fan out of balance when started up. Cleaning and balancing cannot proceed until Unit 11 is shutdown because quillotine isolation damper is inoperable.		22
23			23
24			24
25			25
26			26
27			27
28			28
29			29
30			30

Chart 1 (k)

NIPSCO FGD PLANT
OPERATING HOURS

JULY 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Waiting for Unit 11		1
2	Outage to clean and		2
3	balance booster fan.		3
4			4
5			5
6		11:08 AM Induced draft fan	6
7		repair, booster fan	7
8		cleaning, guillotine	8
9		damper repair, pre-	9
10		cipitator repair.	10
11	Balanced Booster Fan.		11
12	Booster fan and absorber		12
13	started up. Evaporator		13
14	starting up.		14
15	Booster fan at minimum		15
16	speed. Accessing lube		16
17	oil leak on fan bearing.		17
18	Booster fan held at		18
19	minimum speed awaiting		19
20	delivery of oil seals.		20
21			21
22	Repaired oil leak on		22
23	booster fan outboard		23
24	bearing.		24
25			25
26	Repair main steam press		26
27	reducing valve.		27
28			28
29			29
30	Booster fan trip.		30
31	Pressure of main steam		31
	supply to FGD fluctuating		
	down to 280PSIG		
	Repair main steam		
	pressure reducing valve.		
	Booster fan turbine		
	governor repairs.		
	Integrated operation		
	started.		

Chart 1 (1)

NIPSCO FGD PLANT
OPERATING HOURS

AUGUST 1978				
DATE	FGD PLANT		UNIT NO. 11	DATE
1		Integrated Operation.		1
2				2
3				3
4				4
5				5
6				6
7				7
8				8
9				9
10				10
11				11
12				12
13				13
14				14
15	7:20 PM 9:25 PM	Reduction Area Down Due to power failure caused by Electrical Storm.		15
16				16
17				17
18				18
19				19
20				20
21				21
22				22
23				23
24				24
25				25
26				26
27				27
28				28
29				29
30				30
31				31

Chart 1 (m)

NIPSCO FGD PLANT
OPERATING HOURS

SEPTEMBER 1978				
DATE	FGD PLANT		UNIT NO. 11	DATE
1				1
2				2
3				3
4				4
5				5
6				6
7				7
8				8
9				9
10	10:00 AM	Reduction area shut-		
	12:15 PM	down due to SO ₂ compres-		10
11		sor malfunction.		11
12	8:55 AM	FGD Plant shutdown		
		because of boiler	10:41 PM	Scheduled shutdown
13		shutdown. Major		for general maint-
14		FGD Plant related		enance.
		tasks are:		
15		Duct insulation.		Included in the work
16		Booster fan steam		to be performed is
		soot blower.		overhaul of the Unit 11
17		Manual Booster Fan		Air Preheater seals to
		Dampers.		reduce leakage and re-
18		Electric Drive on		moval of intermediate
19		Evaporator Circulating		baskets to effect an
		Pump. Booster Fan		increase in the outlet
20		Turbine Enclosure.		flue gas temperature.
		Rubber Repairs on		
21		Absorber. Sulfur		New pumps for supply of
22		Condenser Tube Maint-		BFW to the FGD system
		enance.		could not be procured
23				in time for installation
				on this turnaround.
24				24
25				25
26				26
27				27
28				28
29				29
30				30

Chart 1 (n)

NIPSCO FGD PLANT
OPERATING HOURS

OCTOBER 1978			
DATE	FGD PLANT	UNIT NO. 11	DATE
1	Annual scheduled boiler maintenance	Annual scheduled boiler maintenance	1
2		continued from last month	2
3			3
4		5:13 PM	4
5		1:16 PM Leak at weld on valve	5
6		9:57 AM Between steam drum and pressure transmitter	6
7	FGD kept down for boiler baseline test by TRW.		7
8			8
9			9
10			10
11			11
12			12
13	Evaporator start-up absorber in operation		13
14	when booster fan was in operation		14
15			15
16	10:30AM Booster fan balanced.		16
17	Louver bypass dampers jammed.		17
18	Booster fan oil cooler problem.		18
19	SO ₂ reduction start-up.		19
20			20
21			21
22			22
23			23
24			24
25			25
26	10:45AM 3:00PM Booster fan governor problem.		26
27			27
28			28
29			29
30			30
31			31

Chart 1 (o)

NIPSCO FGD PLANT
OPERATING HOURS

NOVEMBER 1978				
DATE	FGD PLANT		UNIT NO. 11	DATE
1				1
2				2
3				3
4				4
5				5
6				6
7	7:00 AM	Booster blower shut-		7
8	11:30 AM	down for repair of		8
9		leaking flue gas duct.		9
10				10
11				11
12				12
13				13
14				14
15				15
16				16
17				17
18				18
19				19
20				20
21				21
22				22
23				23
24	2:15 PM	Repair drain valve in		24
25	3:20 PM	Reduction area		25
26				26
27				27
28	10:25 PM	SO ₂ reduction area shutdown		28
29	1:25 AM	due to turbine outage.	10:25 PM	Replace modulating servo
30	12:35 PM	Reduction area down due	1:04 AM	on turbine intercept valve.
	3:30 PM	to variable steam pressure		29
				30

Chart 1 (p)

NIPSCO FGD PLANT OPERATING HOURS

DECEMBER 1978				
DATE	FGD PLANT		UNIT NO. 11	
1	1:30 PM 3:45 PM	Repair evaporator		1
2		heat exchanger tube leaks.		2
3				3
4	5:40 AM 6:00 AM	Reduction section down to low steam pressure.		4
5	10:59 PM	Unit 11 shutdown.	10:59 PM	5
6	2:00 AM		12:49 AM	6
7	10:50 AM 2:40 PM	Reduction section down due to low steam pressure.	Replace quick closing Servo on turbine intercept valve.	7
8	8:35 AM 11:30 AM	Reduction section down.		8
9				9
10				10
11				11
12	9:12 AM	Evaporator fouling due to scaling. Cleaning heat exchanger tubes.		12
13				13
14			8:20 PM	14
15			Clean Air Preheaters.	15
16				16
17			6:54 AM	17
18		Leak in 1st sulfur condenser.		18
19	8:05 AM 11:45 PM			19
20		Tube leaks in evaporator heater.		20
21				21
22	1:06 PM	Absorber on at 8:00 AM		22
23				23
24				24
25				25
26				26
27				27
28				28
29				29
30				30
31				31

$$\text{O.F.} = \frac{\text{HRS. FGD INTEGRATED OPERATION}}{\text{HRS. GENERATOR ON LINE}} \times 100$$

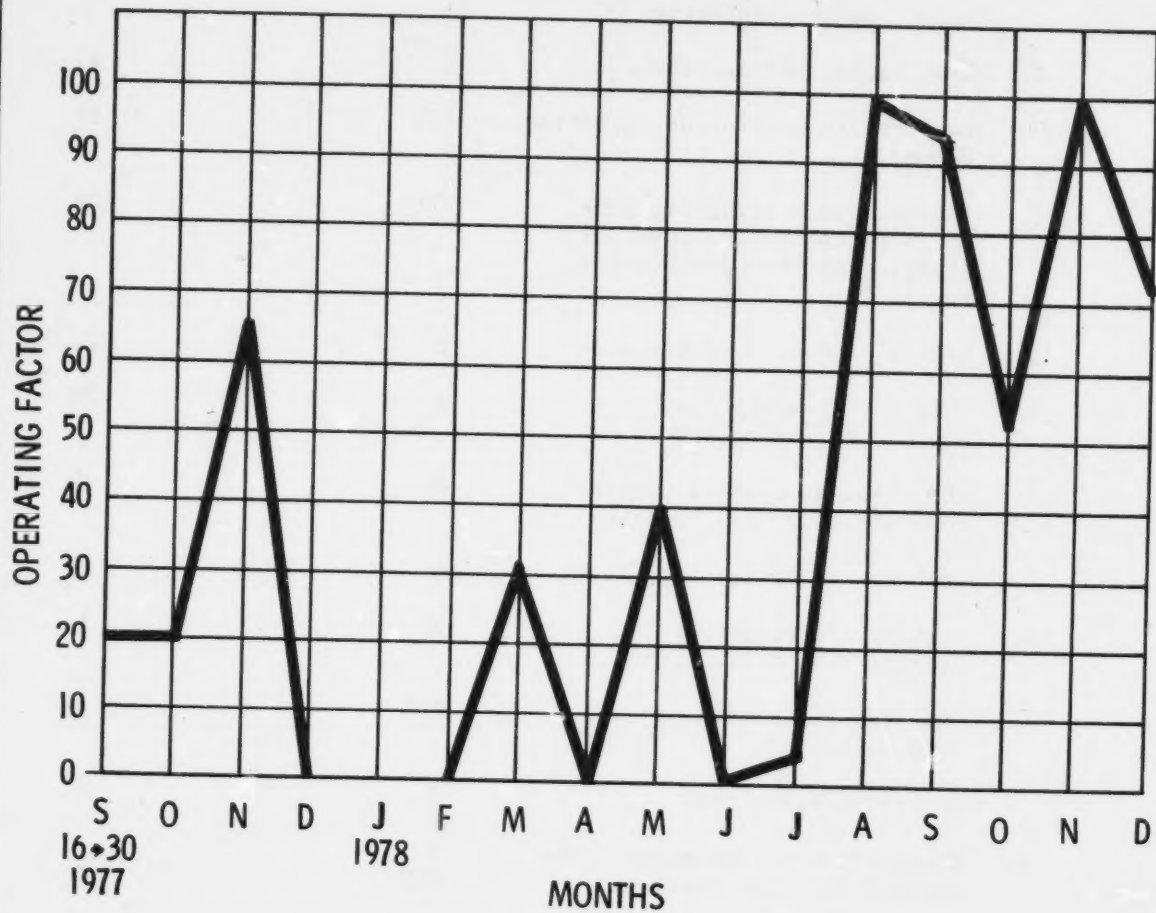


CHART 2
FGD PLANT OPERATING FACTOR (O.F.)
 NIPSCO

TABLE 1

SUMMARY OF FGD OUTAGE CAUSES

Cause	Attributed to Boiler (B), FGD plant (FGD) or combina- tion of both (C)	Total Number of Days in Outage
1. Booster fan cleaning balancing and reblading	FGD	67
2. High boiler silica levels	C(*)	53
3. Booster fan guillotine isolation damper.	FGD	32
4. Booster fan bearing oil leak, booster fan turbine governor repairs, and turbine bearing repair.	FGD	26
5. Scheduled boiler maintenance	B	22
6. Wet, poor quality coal - erratic boiler operation.	B	20
7. FGD plant main steam pressure reducing valve malfunction.	B	13
8. Unit No. 11 boiler tube leaks	B	12
9. Unit No. 11 electrostatic precipitator malfunction.	B	7
10. Evaporator circulating pump repacking.	FGD	7
11. Boiler baseline test	C	6
12. Evaporator heat exchanger tube sealing and tube leaks	FGD	6
13. Heat balance and flue gas flow rate tests.	B	6
14. Bearing failure - evaporator circulating pump turbine drive.	FGD	5
15. Induced draft fan imbalance	B	5

<u>Cause</u>	<u>Attributable to Boiler (B), FGD plant (FGD) or combina- tion of both (C)</u>	<u>Total Number of Days in Outage</u>
16. Coalescer pluggage.	FGD	5
17. Gasket failure - evaporator solution line.	FGD	4
18. Turbine-generator stop valve repair.	B	4
19. Air preheater cleaning	B	3
20. Various instrument problems	FGD	3
21. Tailgas incinerator malfunction.	FGD	2
22. Sulfur condenser tube leaks.	FGD	2
23. SO ₂ compressor gasket leak.	FGD	2
24. Process water booster pump failure	FGD	2
25. Steam leak at drum pressure transmitter line.	B	1

RECAPITULATION

Total days in period September 16, 1977 through December 31, 1978.	472
Total days integrated FGD operation.	157
Total days FGD outage.	315
Total days FGD attributable outage.	163
Total days boiler-turbine attributable outage.	80
Total days combination attributable outage.	59
Total days boiler and turbine operated (on line)	424
Total days boiler outage (off line)	48
Longest continuous integrated FGD operation period.	43 days (disregarding 2-2 hour interruptions in reduction section).

MAJOR FGD OUTAGES

	<u>Days</u>	<u>% of Total Outage</u>
Booster fan related	93	30
High silica levels in boiler	53	17
Guillotine damper related	32	10
Coal quality related	20	<u>6</u>
Percent of total outage		63

- (*) The cause of high silica in the boiler feed water was a silica breakthrough in the portable demineralizer unit that was being used to supplement the increased water demand. The additional usage was compounded by increased boiler blowdown. This was corrected by setting more stringent silica limits and closer monitoring equipment of the portable demineralizer unit. It was also followed up by a program to reduce condensate consumption and losses including installation of automatic reset controls on the return condensate diversion system. (See pages 11 and 12, item number 6.)

10. FGD System Imports and Exports

Table 2 on page 740 lists the monthly Imports and Exports of the FGD system covering the time period of September 1977, through December 1978.

	SEPT. 16th-30th 1977	OCT.	NOV.	DEC.	JAN. 1978	FEB.	MAR.	APRIL	MAY	JUNE	JULY	AUG.	SEPT.	OCT.	NOV.	DEC.		
IMPORTS																		
ELECTRIC POWER (1000 KWH's)	223	417	452	434	412	368	511	253	472	412	426	536	377	385	517	497		
STEAM (1000 lbs.) (*)	21301	29433	34302	28428	15544	17142	34420	12809	35952	21551	28422	47683	17513	25145	40888	37511		
NATURAL GAS (1000 SCF)	2487	4074	5526	3333	3307	2707	3732	1885	3946	1807	2052	7863	3250	3671	6822	5106		
CONDENSATE MAKE UP (1000)	498	14067	16443	16533	12495	6561	14661	262	8422	6084	7240	14094	6633	6944	10907	14409		
POTABLE WATER (1000 lbs.)	2054	3568	2384	4791	1299	778	2639	1252	5111	4525	4488	5665	5191	5554	6055	3178		
BOILER FEEDWATER (1000 lbs.)	584	2288	1749	1730	1415	955	1750	1529	2404	2288	2957	2539	2484	2417	2392	2555		
SODA ASH (NET TONS)	0	67	171	97	0	12	235	15	251	53	108	282	103	105	83	83		
FLUE GAS (AVE. PPMV SO ₂)	2020	2287	2373	-	-	-	2016	-	1823	2243	1797	2182	2221	2283	2190	2044		
COOLING WATER							DESIGN FLOW RATE = 4,545 GPM											
EXPORTS																		
SULFUR (LONG TONS)	0	102	285	0	0	0	135	0	191	0	18	526	171	227	513	305		
SODIUM SULFATE PURGE (NET TONS)	0	14	56	12	29	28	41	14	44	45	26	65	33	36	68	9		
CONDENSATE RETURN (1000 lbs.)	10200	30888	18157	9787	4486	764	24642	10554	26126	14337	20440	30323	17073	35042	43256	36445		
FLUE GAS (AVE PPMV SO ₂)	188	230	228	-	-	-	230	-	205	288	198	225	229	250	217	220		
FLY ASH PURGE							16 to 50 GPM, 1% SOLIDS OR LESS											
OTHER DATA																		
FGD PLANT INTEGRATED OPERATION (HRS)	67	131	428	0	0	0	215	0	263	3	30	742	271	324	713	479		
UNIT II OPERATION-ON LINE (HRS)	371	709	660	665	654	478	690	544	682	720	658	774	287	634	717	682		
AVERAGE SO ₂ REMOVAL EFFICIENCY (%)	90.7	89.9	90.4	89.8	-	-	88.5	-	88.3	87.1	88.8	89.7	89.6	89.0	90.1	89.1		
UNIT II NET GENERATION (1000KWH's)	49549	44534	44534	42101	50498	40208	43270	37168	48076	49429	39698	51860	19719	50878	49879	48999		
AVE. NET MWE	51	70	67	63	77	84	63	68	70	69	61	70	69	80	70	71		

(*) STEAM/CONDENSATE IMBALANCES ARE DUE TO: 1. METERING INACCURACIES.

2. EMERGENCY STEAM NOT METERED UNTIL 11-78.

3. PORTION OF CONDENSATE MAKE UP TO UNIT II BOILER IS ESTIMATED.

TABLE II

NORTHERN INDIANA PUBLIC SERVICE COMPANY
DEAN H. MITCHELL STATION - UNIT NUMBER II
FGD PLANT IMPORTS AND EXPORTS

11. FGD System Operating and Maintenance Costs

Table 3 on page 742 is a summary of FGD Operating, Maintenance and Improvement Costs for September 16, 1977, through December 31, 1978.

TABLE 3

OPERATING & MAINTENANCE COSTS

NIPSCO UNIT NO. 11 FGD PLANT

Operating, Maintenance and Improvement Costs are listed from the period of September 16, 1977, through December 31, 1978.

Operation and Maintenance - Offsites Facilities (including booster blower, flue gas ductwork and dampers, utilities system)	\$ 520,700 (a)
Operation and Maintenance - FGD Process (Includes by-products storage and loading, raw materials unloading and storage)	3,309,200 (b)

UTILITIES

Steam @ \$2.00/1000 lb	895,500
Demineralized water September 1977 through February 1978 (\$0.03/gallon) March 1978 through December 1978 (\$0.0125/gallon)	531,900 (c)
Electric power @ \$0.024/kWh	160,600
Natural gas @ \$1.9812/million Btu	<u>121,900</u>
Total Utilities	\$1,709,900
Total FGD Costs before by-product credit	\$5,539,800
Credit for sulfur and sodium sulfate	<u>(97,000)</u>
Total FGD Operating, Maintenance and Improvement Costs after by-product credit	\$5,442,800

(a) Includes \$20,000 for installation of manual slide gates at booster fan inlets.

(b) Includes approximately \$150,000 for one time mechanical system modifications.

(c) Water costs were abnormally high due to use of demineralized water as condensate for flushing pump seals at 0.03 per gallon versus use of their own process water at an estimated cost of \$.12 per one thousand gallons. Approximately 75% of the condensate consumption in the FGD system is attributable to pump seal flushing.

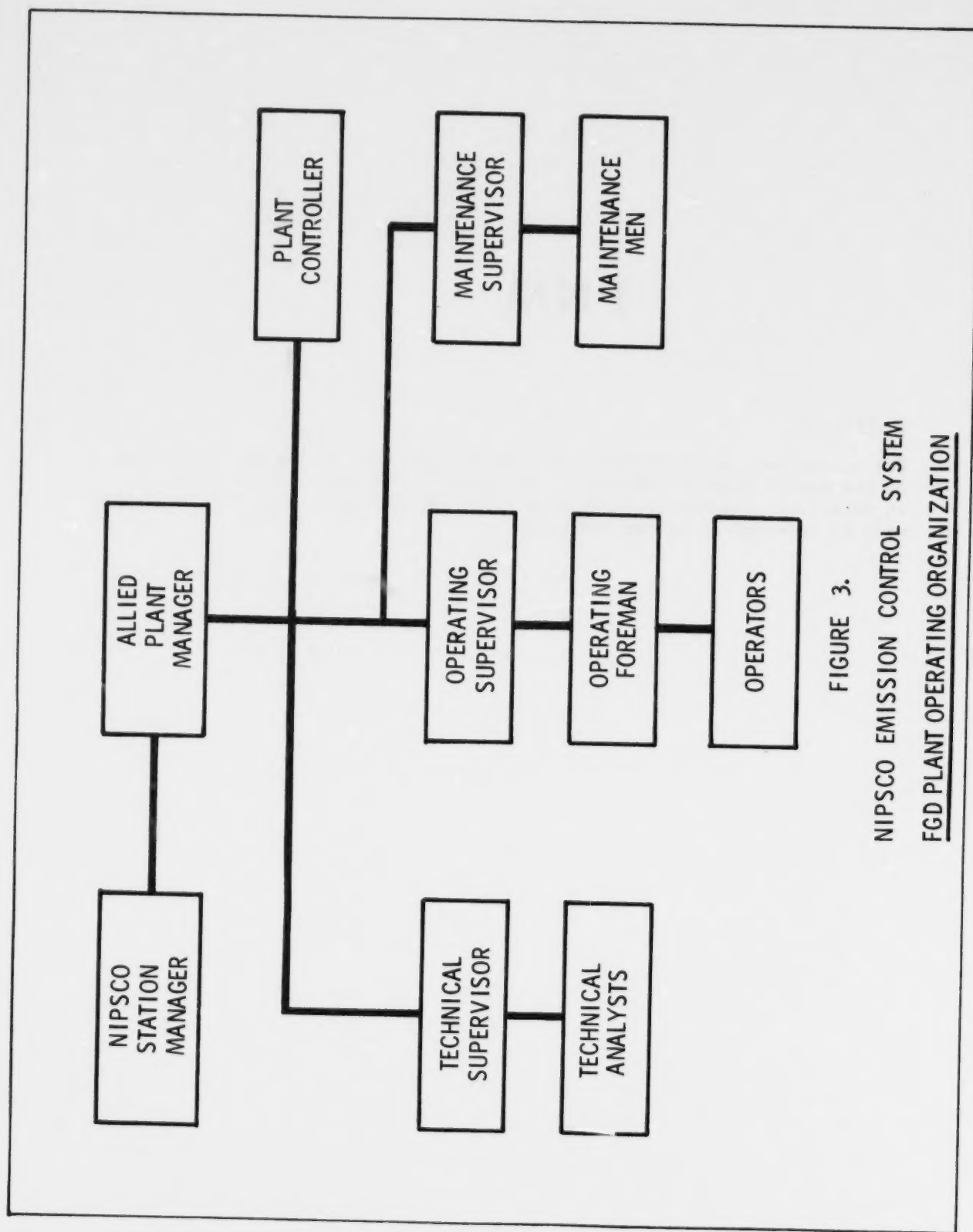


FIGURE 3.

NIPSCO EMISSION CONTROL SYSTEM
FGD PLANT OPERATING ORGANIZATION

PNM

Section II

The second part of this paper, Section II, pertains to the FGD facilities of the Public Service Company of New Mexico's San Juan Station located in Waterflow, New Mexico (near Farmington, New Mexico). The FGD facilities will be referred to as the PNM plant.

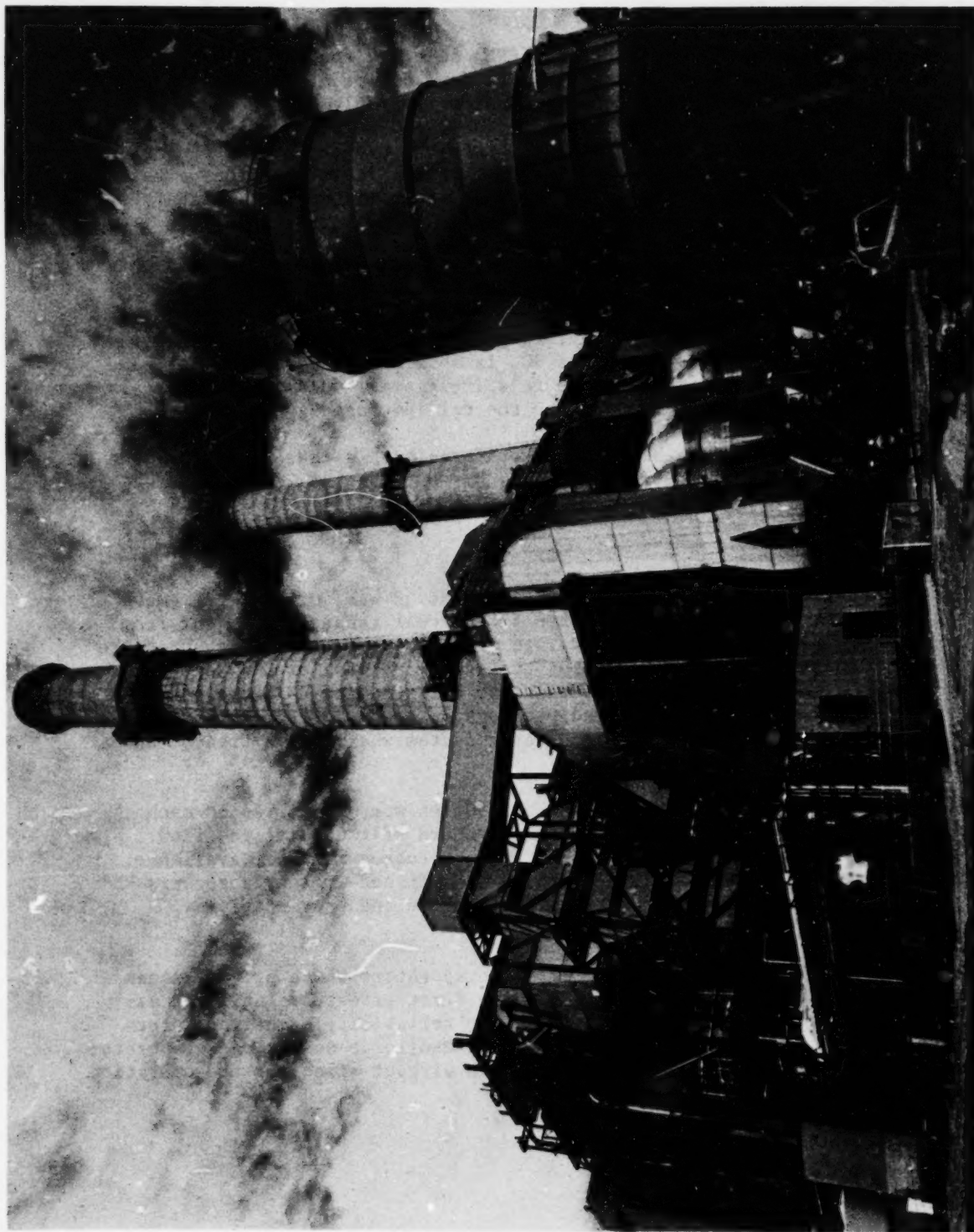


Figure 4
GENERAL VIEW OF PNM PLANT

1. PNM Project Background

Public Service Company of New Mexico (PNM) selected the Wellman-Lord/Allied Chemical FGD processes early in 1974 from four different systems that were being considered. The process compared favorably both economically and technically with the lime-limestone, double alkali, and with a dry char adsorber. The Wellman-Lord/Allied Chemical process was also considered advantageous because the end product, elemental sulfur, could be marketed. The initial concept was to regenerate the purge salt, sodium sulfate. In the meantime, a contract had been obtained to sell this material. The elimination of the sulfate regeneration process reduced the capital requirements for Units No. 1 and 2 by about \$4,000,000.

A major advantage of the Wellman-Lord process is using a clear scrubbing solution that prevents absorber pluggage. A more soluble substance is produced in the scrubbing liquor as SO_2 is absorbed. The materials handling is also much simpler since relatively small volumes of materials enter and exit the system compared to the calcium based system.

The main disadvantage of the Wellman-Lord process, is that it is somewhat more complex than the lime-limestone systems because the solutions are regenerated.

2. PNM Design Criteria

The Wellman-Lord system for Units No. 1 and No. 2 was designed to remove 90% of the SO_2 from the flue gas when firing coal ranging in sulfur from 0.59% to 1.3% by weight with an average of 0.8%. Relatively high pressure drop prescrubbers were specified for the system because of New Mexico's stringent regulation of a fine particulate emission and to provide some back up for the electrostatic precipitators. The purge from the prescrubber is sent to the plant waste water system where it is treated and then recycled.

At PNM's request four scrubber absorber modules were installed on each power plant unit each sized to handle one-third of the total gas flow. Therefore, the plant has one complete spare scrubber-absorber available. This permits a maintenance program to be established whereby absorbers can be rotated in and out of service for routine and preventative maintenance purposes.

The chemical plant has two double effect evaporators which provide steam conservation (the overhead vapors from the first effect are utilized as the heat source in the second effect). For reliability each evaporator is connected to steam and offgas compressor manifolds so any one evaporator can be taken out of service for maintenance without affecting the operation of the three absorber units.

The purge treatment plant consists of three low temperature crystallizers where sodium sulfate is precipitated in the decahydrate form. The crystallizers are followed by a melt tank and evaporator. The evaporator is similar to, but smaller than, the main evaporators. Water is driven off and the sulfate purge is centrifuged, then dried in a flash dryer. The entire purge treatment plant was specified to obtain a concentration of 70% sodium sulfate and 30% sodium sulfite in the dried purge salt. The actual sulfate content has been very high; purities have been achieved in the area of 90% sodium sulfate. Residual moisture in the dried purge salt has been less than 1.0%.

Two identical Allied Chemical SO_2 reduction trains were installed as part of Units 1 and 2 FGD systems. Each of the trains has a design capacity of more than 50% of the total FGD system capacity based on the use of low grade coal (1.3% sulfur) in the boilers. A requirement for two (2) SO_2 reduction trains was specified by PNM with the objective of achieving an FGD system with essentially a 100% on-stream reliability.

The system being designed for the Unit Number 3 and 4 power plants will be somewhat similar with the following exceptions. The prescrubbers on Unit Number 3 and 4 system will have a lower pressure drop for energy considerations. The five stage tray absorbers also function quite well for residual particulate removal; sulfate purge quality will not be degraded since the fly ash is filtered out of the solution.

In the interest of economy the absorbers have been designed for 4 units in operation per boiler when burning low grade coal; i.e., this removes the one module spare as on Units 1 and 2. The coal being used at San Juan Station has rarely exceeded 0.95% sulfur for long durations of time hence this should not decrease the plant operability and maintenance capability.

A sulfuric acid unit will be installed in the FGD system for Units 3 and 4. The sulfuric acid plant design capacity will be based on low grade coal (470 tons per day as 100% sulfuric acid). The rationale for selection of sulfuric acid rather than sulfur as the by-product is presented in a later portion of this paper.

3. Plant Operations

General - The plant scrubber operation is controlled by two separate PNM groups; the power generation plant personnel are responsible for the booster blowers, scrubbers, absorbers, fly ash filters, and waste water treatment. The chemical plant operations personnel are responsible for supplying solutions for the absorbers, regeneration of SO_2 from the absorbing solutions, operation of the sulfur dioxide reduction unit, purge treatment, and loading and shipping of the products. Maintenance people at San Juan are organized by zones. Figures 5 and 6 illustrate the organization structure for operations and maintenance.

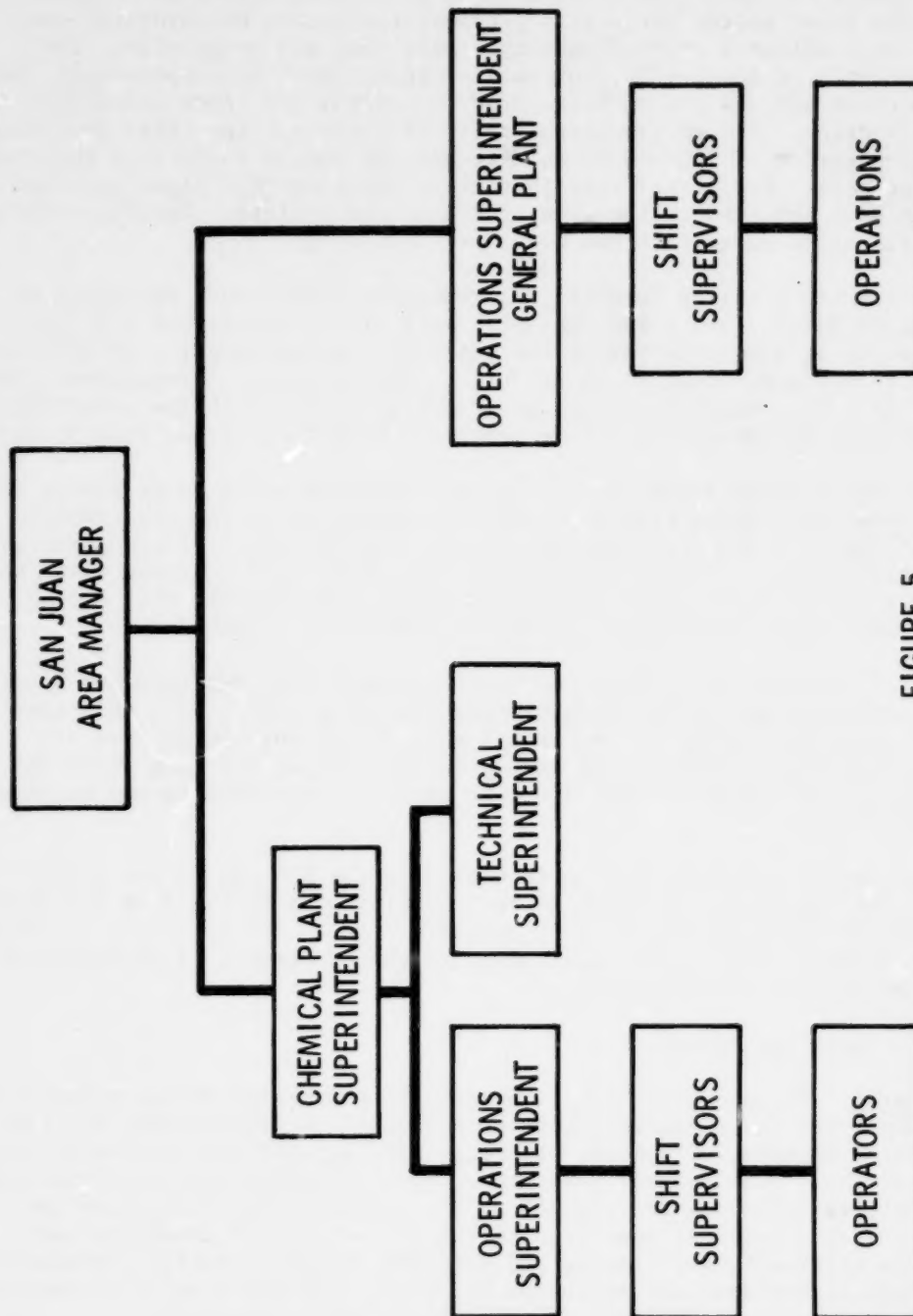


FIGURE 5.

PNM OPERATIONS ORGANIZATION

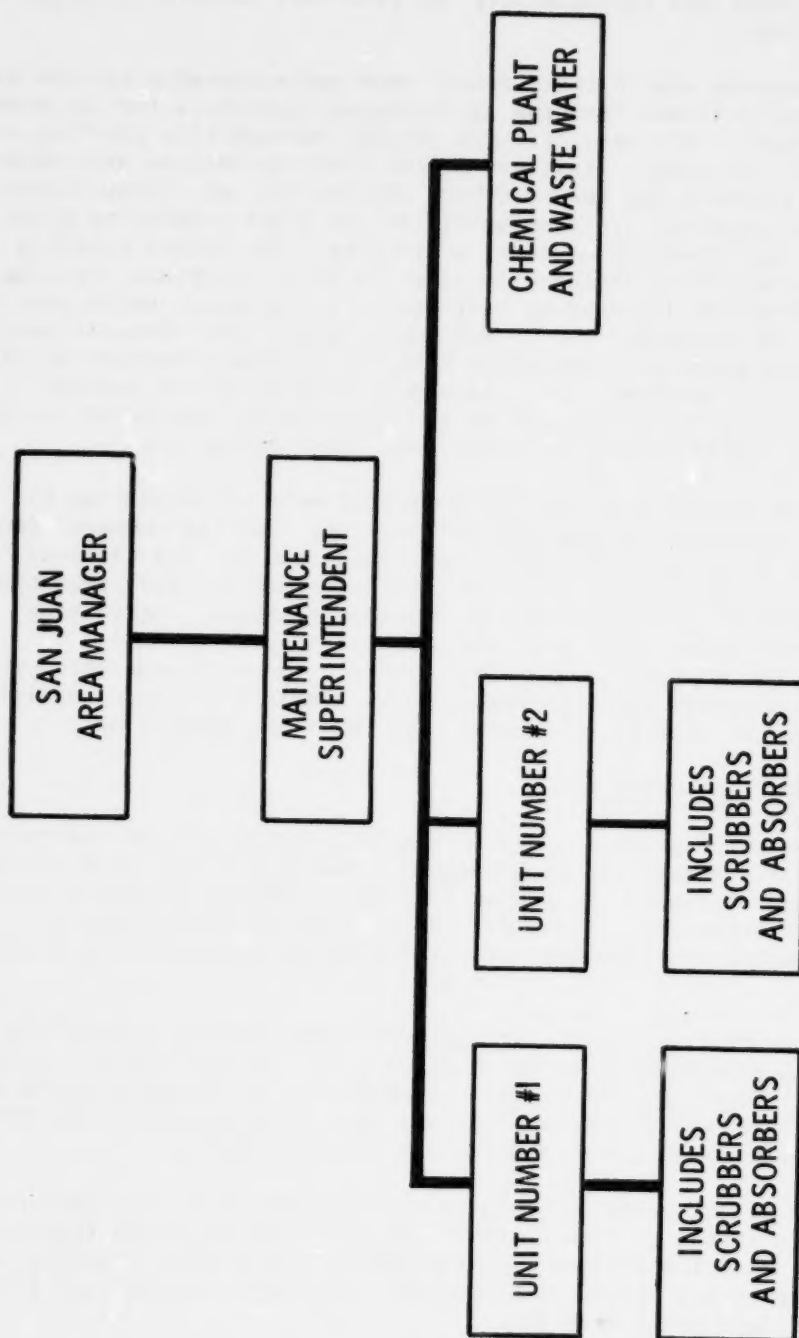


FIGURE 6.

PNM MAINTENANCE ORGANIZATION

Personnel - There are approximately 100 personnel involved in support of the FGD process.

Scrubber operators are in their fourth year apprenticeship and are near the end of their formal training as Journeymen Operators for the power generation plant. All operators are rotated through this position as part of their training. First and second year apprentices are assigned the field activities for the scrubbers and the fly ash filters areas. All of these operators are responsible to the power generation plant supervision for direction of their activities. The formal training given to the operators included lectures by Davy. In-plant training during start-up was directed by Davy start-up engineers, while some of the "on the job training" was by PNM supervision. The chemical plant operators were given more extensive training initially because of the chemical process involved. This included training during initial equipment testing for acceptance by the engineering contractor, as well as the plant commissioning and subsequent plant operations.

Approximately one-third of the FGD personnel were selected from the power plant operations group and the remainder from the chemical industry. The creation of an operationally experienced crew for the chemical plant, with the necessary levels of experience and ability presented numerous problems of finding and relocating personnel. Recruiting outside of the local plant area was necessary because of local competition for experienced people. Past experience of operators and supervisors included ammonia plants, power generation plants, sulfur plants, chemical plants, refineries, and PNM's own power plant.

4. How The Plant Operates

Scrubber-absorber operations does not affect the power plant operation because the FGD system can be by-passed. During a normal unit start-up, scrubbers and absorbers are put on line after the electrostatic precipitators are functioning. Power for the electrostatic precipitators, booster blowers, and absorber circulating pumps is supplied from PNM's power plant.

Each scrubber-absorber module is operated independently (except the control data transmission to the control panel board), and put on line separately. Each unit has a reheat system that is needed to protect the stack from corrosive products resulting from condensation in the flue gas when three modules are in operation simultaneously.

Two 750,000 gallon tanks for absorber product and feed solution provide surge in the system to prevent the chemical plant operation from being affected by normal operating fluctuations in the scrubber-absorber area. The ideal operating situation is to have the feed solution tank full and the product tank level very low.

The chemical plants ability to operate depends on the power generation unit operations. To date this has been the major operating problem. Steam and water to the chemical plant originates in the power plant area; clean condensate and process offgas are returned from the chemical plant. The sumps and waste water streams are collected and neutralized in a water treatment plant.

SO₂ is recovered from the absorber product solution in a double effect evaporator. Precise control of the rate of SO₂ recovery from the slurry in the evaporators is possible. The oxygen level in the recovered SO₂ stream is controlled at less than 0.35%, as per the design which facilitates operation of the SO₂ reduction units at maximum natural gas utilization. Evaporator slurry concentrations up to 70% by volume have been achieved with little or no solids accumulation in any of the vessels. Condensed water from the overhead SO₂-H₂O stream is recycled and used to dissolve the solids from the evaporators. Both evaporator trains have operated for many consecutive days without interruptions (other than steam availability) over a wide range of operating conditions. Critical parameters for SO₂ recovery rate are: steaming rates, slurry solids percent, and chemical composition. Enough versatility has been provided so that all four evaporators may be operated simultaneously or independently.

High speed, dry compressors are used to lower the absolute pressure in the evaporators and pressurize the SO₂ gas stream through the reduction process. Offgas blowers downstream of the SO₂ reduction units serve to return incinerated tail gas and vessel vent gases back to the scrubber-absorbers.

Each of the Allied Chemical SO₂ Reduction units includes a primary reactor system where a portion of the SO₂ is reduced to sulfur and H₂S using natural gas as a reductant. Exothermic heat of reaction is stored in combination reactor generator vessels for subsequent use in preheating the feed gas. The flow through the primary reactor system is reversed on a periodic basis. The sulfur formed is condensed and the cooled gas, containing proper proportions of SO₂ and H₂S, is processed through a Claus conversion system for recovery of additional sulfur. The Claus system off-gas is incinerated and recycled to the Wellman-Lord absorbers.

The raw natural gas available at the San Juan Station contains quantities of C₅ and higher hydrocarbons which make it unacceptable for use as a reductant in the SO₂ reduction process. The heavy ends are removed in a gas treating unit and utilized as fuel in the tail gas incinerators.

Sulfate formed by the the oxidation of sulfite in the absorber solution must be purged from the solution since sulfate is inert in the system and will not absorb or release SO₂. Sulfate is separated from the absorber product solution by low temperature crystallization.

The sulfate is recovered as a decahydrate, melted and sent to an evaporator for removal of the water of hydration, then separated and dried to less than 1% water for ease of storage and shipment.

Separate plant utilities are dedicated to the chemical plant for cooling towers, air compressors, steam reducing stations, water supply and recovery systems. Steam, high quality water, and electricity as stated previously are provided from the PNM generating plant. Steam condensate is returned to the generation plant from the Chemical Plant.

5. Plant Operations

While considerable time will be spent on discussing operational problems, it is worth mentioning those units that have operated as designed and with little or no start-up problems.

Computer control of the scrubbers from the remote control room has worked well from the beginning. As with most computer-controlled systems, the success or failure of the control system depends on the sensor elements in the field.

Absorbers do a good job of removing SO_2 from the flue gases. With nearly design solutions and normal amounts of soda ash being added to the system, Unit Number 2 has made compliance with State and Federal regulations as demonstrated on November 29, 1978. Unit Number 1 compliance will be demonstrated as soon as possible.

Evaporators and SO_2 compressors have been operated for long periods of time. The only problem has been some erosion of the SO_2 compressor impellers probably caused by a condensing condition at the inlet of the compressors. Corrections and revisions have been made to prevent this condensation. These units have been operated, since the corrections, over a wide range of flow rates with no difficulty.

Soda ash addition to the process has been relatively easy and very dependable. Soda ash is stored dry and dissolved when needed for use at each absorber or at the evaporator slurry dissolving tank.

6. Plant Operating Problems

As in most new plants there were various mechanical problems experienced during the initial start-up. At times these problems created operating difficulties. Most of these problems have been resolved with a few to be completely cleared. The following is a brief summary of the major problems that have occurred.

Steam and Water Availability:

Problem:

The steam and water supply to the FGD plant has frequently been interrupted or curtailed thereby greatly reducing or stopping the SO_2 chemical plant operations. This resulted from operating difficulties in power generation units.

Solution:

The original design parameter was that relatively unlimited steam and water supplies would be available from the power plant. This has not been true. Studies and evaluations of the steam and water systems are underway to assure more continuous supply, or to provide standby quantities of steam and boiler quality water solely dedicated to the chemical plant.

Results:

Not enough time has elapsed for results to be evaluated pertaining to changes or modifications that have been made or completed.

Solution Losses:

Problem:

There have been solution losses occurring which have not been readily identifiable, especially in the winter months. This solution loss appears as though it may be coming from the bottom tray of the absorber into the scrubber sump when less than minimum design gas flow rate is being processed.

The design turndown is 50% for each absorber. The solution losses occur when the absorbers are operated below the design turndown in order to limit the amount of incoming flyash during periods when the electrostatic precipitators malfunctioned. Additional solution losses have occurred at the fly ash filter whenever the automatic cycling fails.

Solution:

- a. Improved operator attention.
- b. Correction of the electrostatic precipitator problems
- c. Operate the absorbers at or above the minimum design gas flow rate.

Results:

Not all of the planned corrections have been completed or performed because of scheduling and operation of the plant. Plant turnaround in the future during which these changes are to be accomplished.

Electrostatic Precipitator:

Problem:

The electrostatic precipitators have failed or malfunctioned leading a condition which overloads the scrubber solution with fly ash. Design specifications are that the scrubber should be capable of accepting a complete precipitator failure for a period of two (2) hours. The

scrubbers have demonstrated the capability of accepting a complete precipitator failure for time periods up to 36 hours. Furthermore, the scrubbers have operated for extended periods of time (several days) when the precipitators operated at less than 60% efficiency. Solution density measurement devices have proved inadequate to warn of pending recirculation problems. At one time significant amounts of fly ash were collected in the recirculating solution to increase the density where recirculation was reduced to allow hot flue gases to bypass. The hot gases contacted the down stream heat sensitive Chevron mist eliminators that separate the scrubber and absorber. Some of the Chevrons were warped and required replacement in four modules.

Solution:

- a. Correction of problems in the electrostatic precipitators.
- b. Improved operator attention and manual sampling.
- c. Establishment of operating techniques to handle sudden and unexpected fly ash, such as setting a 15% by volume fly ash concentration limit for shut down.

Results:

No real results will be available until all the planned improvements have been executed. These corrections are to be performed during the next planned plant turnaround scheduled in the near future.

Purge System:

Problem:

The purge system evaporation of crystallized and remelted sulfate to a dry solid has not operated sufficiently to provide evaluation. This is the result of power plant shutdowns and steam and water shortages. There has been some indication that solids being formed in the purge evaporator are too small to be separated by a screen type centrifuge. A problem has also been experienced in the method of feeding the wet centrifuge cake into the flash dryer.

Solution:

- a. Improved utilities, such as steam, water and power.
- b. Sieve sized changes in the centrifuge.
- c. Improved operating techniques for improved crystal growth.
- d. Improvement in conveying chutes, conveyor and ducts.

Results:

To date results have been difficult to evaluate because of insufficient operating time.

Claus Catalyst Beds:

Problems:

Portion of catalyst in the SO₂ reduction area was found to have shifted past the Claus converter catalyst support screens into the bottom of the vessels. Investigation revealed that support screens had been improperly installed during plant construction with substantial gaps between sections of screen and adjacent to the vessel walls.

Solution:

Correct installation of the screens in the Claus converter has been completed.

Results:

The problem has not reoccurred since correction was made.

7. Future Plans & Chemical Sales Program

The sales of FGD process products, the main concern in the immediate future, has several effects on the operating and the administration costs of the scrubber operations. Since beginning the chemical sales contracts, PNM has received numerous contacts concerning contracts and logistics involved in selling and shipping these materials. The following is a brief discussion that attempts to answer the typical inquiries received on this program. The sales of these commodities have been executed through a broker.

The most obvious cost reduction effected by the sales program is the elimination of disposal costs. On an average basis the four units (two future units to be built) at San Juan Station will produce about 300 tons/day of SO₂. This translates into 150 tons per day of elemental sulfur or 1340 tons per day of 60% calcium sulfate and water sludge had limestone scrubbers been used. Whereas the sulfur produced in a regenerable system is a salable product which results in a significant monetary credit to help offset the operating cost of the FGD system, use of a limestone scrubber FGD system would necessitate the disposal of some 45 truck loads of sludge per day. Gypsum disposal costs for the plant would have exceeded \$1,000,000 per year for hauling, mine preparation costs excluded.

PNM plans to install a sulfuric acid plant on the FGD Units Number 3 and 4, which is presently in the engineering phase. The advantage of producing sulfuric acid instead of sulfur is the elimination of natural gas consumption. At \$1.70/MCF Units Number 1 and 2 consume about \$450,000 per year of natural gas. With sulfur plants on Units Number 3 and 4, San Juan Station would require \$1,007,000 per year of natural gas. A sulfuric acid plant costs less in capital than a comparable capacity SO₂ reduction unit. Although the natural gas is classed as chemical plant feed stock, it is still subject to curtailment during the winter months, which is another reason to eliminate this requirement.

When the San Juan complex is complete the following materials will be produced from the scrubber systems. (On an average coal basis):

<u>Item</u>	<u>Approximate Short tons per Day</u>	<u>Truck loads per Day</u>
1. Salt Cake	60	2
2. Sulfur	60	2
3. Sulfuric Acid	250	10

TABLE 4

Anticipated PNM Material Produced

While it would seem inappropriate to discuss unit prices here, it can be stated that the revenues PNM will receive for these materials will reduce the overall FGD system operating costs by about 10%, or roughly \$1,000,000 per year, in addition to the savings discussed earlier.

Revenues received by others for product chemicals depends on site location and the availability of transportation. The net return on chemicals is freight sensitive and distances of 100 miles can easily double the selling price if one expects to recover the freight. Easy access to rail or barge networks would allow cheaper freight, hence the radius encompassing final customers is extended somewhat if those options are available. At times these commodities may have to be sold on a freight equalized basis; that is, the seller absorbs part of the freight in order to offer a competitive price for the material. The extent of freight equalization depends upon market conditions, the proximity of other production plants, and upon what modes of transportation are available at the plant site. In any case, it is possible that freight costs can reduce the possible return on the product sales.

Since the market prices for sulfur and sulfur products have varied considerably over the years, the contract between a broker and a producer should be structured to protect both parties. One form is the standard "evergreen" contract with a floor price for the producer with an additional increment for the broker to cover his costs. Anything beyond the increment could be split on a percentage basis that is mutually agreeable to the contracting parties.

One aspect of scrubber operations that is affected through sales of products is that the quality of the end product must be considered in daily operations. For salt cake this means color, particle size, and composition. For sulfur, quality is based on color, purity, and the absence of remelt sulfur. The sulfuric acid quality will be clear (free of particulates) and have a low iron content. The Wellman-Lord scrubber system is expected to produce electrolytic grade acid because the SO₂ from the evaporators is very clean.

<u>COMPANY AND LOCATION</u>	<u>FEED GAS ORIGIN</u>	<u>SCFM GAS TREATED</u>	<u>DISPOSITION OF SO₂</u>
<u>UNITS ON STREAM</u>			
Olin Corp./Pauleboro, NJ (now shut down)	Sulfuric Acid Plant	45,000	Recycle to Acid Plant
SOCAL /El Segundo, CA	Claus Plant	30,000	Recycle to Claus Plant
Allied Chem./Calumet, IL	Sulfuric Acid Plant	30,000	Recycle to Acid Plant
Olin Corp./Curtis Bay, MD	Sulfuric Acid Plant	78,000	Recycle to Acid Plant
SOCAL /Richmond, CA	Claus Plant	30,000	Recycle to Claus Plant
SOCAL /Richmond, CA	Claus Plant	30,000	Recycle to Claus Plant
SOCAL /El Segundo, CA	Claus Plant	30,000	Recycle to Claus Plant
NIPSCO/Gary, IN	115 MW Coal Fired Power Plant	310,000	Elemental Sulfur Plant
PSCNM/Waterflow, NM	700 MW Coal Fired Power Plant	1,880,000	Elemental Sulfur Plant
<u>UNITS IN DESIGN OR CONSTRUCTION</u>			
PSCNM/Waterflow, NM	1100 MW Coal Fired Power Plant	2,727,000	Sulfuric Acid Plant
Getty/Delaware City, DE	Three Coke Fired Boilers	500,000	Sulfuric Acid Plant

EXHIBIT I

WELLMAN-LORD PLANT INSTALLATIONS IN THE UNITED STATES

COMPANY AND LOCATION	FEED GAS ORIGIN	SCFM TREATED	DISPOSITION OF SO ₂
UNITS ON STREAM			
Japan Syn. Rubber/Chiba	Oil Fired Boiler	124,000	Sulfuric Acid Plant
Toa Nenryo/Kawasaki	Claus Plant	41,000	Recycle to Claus Plant
Chubu Electric/Nagoya	220 MW Oil Fired Power Plant	390,000	Sulfuric Acid Plant
Japan Syn. Rubber/Yokkaichi	Oil Fired Boiler	280,000	Sulfuric Acid Plant
Sumitomo Chem./Sodegaura	Oil Fired Boiler	225,000	Sulfuric Acid Plant
Kashima Oil/Kashima	Claus Plant	20,000	Recycle to Claus Plant
Mitsubishi Chem./Mitsushima	Oil Fired Boiler	373,000	Sulfuric Acid Plant
Toa Nenryo/Hatsushima	Claus Plant	10,000	Recycle to Claus Plant
Toyo Rayon/Nagoya	Oil Fired Boiler	218,000	Sulfuric Acid Plant
Japan Nat. Railroad/Kawasaki	200 MW Oil Fired Power Plant	435,000	Sulfuric Acid Plant
Kurashiki Rayon/Okayama	Oil Fired Boiler	248,000	Sulfuric Acid Plant
Fuji Film/Fujinomiya	Oil Fired Boiler	89,000	Liquid SO ₂
Shin Daikyowa/Yokkaichi	Oil Fired Boiler	253,000	Sulfuric Acid Plant
Sumitomo Chem./Niigata	Oil Fired Boiler	91,000	Liquid SO ₂
Mitsubishi Chem./Mitsushima	Oil Fired Boiler	390,000	Sulfuric Acid Plant
Mitsubishi Chem./Kurosaki	Oil Fired Boiler	230,000	Sulfuric Acid Plant
Tohoku Electric/Niigata	100 MW Oil Fired Power Plant	236,000	Sulfuric Acid Plant

EXHIBIT 2

WELLMAN-LORD PLANT INSTALLATIONS OVERSEAS

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E. L. Mann
Executive Manager of Electric Production
Northern Indiana Public Service Company
5625 Hohman Ave.
Hammond, Indiana 46325

L. C. McGrath
Manager of Plant Engineering
Northern Indiana Public Service Company
591 Marquette Mall Office Building
490 St. John Road
Michigan City, Indiana 46360

D. W. Christian
SO₂ Chemical Engineer
Public Service Company of New Mexico
Albuquerque, New Mexico 87103

Robert Bierbower
Project Manager
Allied Chemical
Industrial Chemicals Division
P.O. Box 1139R
Morristown, New Jersey 07960

A. Giovanetti
Supervising Process Engineer
Davy Powergas Inc.
P.O. Drawer 5000
Lakeland, Florida 33803

R. I. Pedroso
Principal Process Engineer
Davy Powergas Inc.
P.O. Drawer 5000
Lakeland, Florida 33803

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January 1979
3. Sulphur Recovered From SO₂ Emissions at NIPSCO's Dean H. Mitchell Station
Howard A. Boyer
Allied Chemical Corporation
Morristown, New Jersey

and

Roberto I. Pedroso
Davy Powergas Inc.
Lakeland, Florida
4. "Sulfur Recovered From Flue Gas at Large Coal Fired Power Plants"
Roberto I. Pedroso
Davy Powergas Inc.
Lakeland, Florida

Paper presented to the Third Symposium on Sulphur and other Airborne Emissions, Salford, England - April 1979

**CITRATE PROCESS DEMONSTRATION PLANT
— CONSTRUCTION AND TESTING —**

Authors:

Richard S. Madenburg
Morrison-Knudsen Company, Inc.
Boise, Idaho

Laird Crocker
U.S. Bureau of Mines
Citrate Process Demonstration Plant
Monaca, Pennsylvania

John M. Cigan
St. Joe Zinc Company
Monaca, Pennsylvania

Laurance L. Oden
U.S. Bureau of Mines
Albany Metallurgy Research Center
Albany, Oregon

R. Dean Delleney
Radian Corporation
Austin, Texas

UNITED STATES ENVIRONMENTAL PROTECTION AGENCY

Flue Gas Desulfurization Symposium

Regenerable Processes Session

March, 1979

Abstract

Construction of the Citrate Process Demonstration Plant represents a major milestone in the achievement of the Federal Bureau of Mines goal to minimize the undesirable environmental impact associated with SO₂ emissions from industrial sources. Baseline performance testing of the boiler prior to retrofit of the Citrate Process was conducted during November, 1978. Mechanical testing, start-up and operational commissioning of the plant are now in progress. The demonstration of the Citrate FGD process at a 60 MWe coal-fired power generating station will confirm the design basis, the technical merits and the process's capital and operational economics at a commercial scale installation.

This paper discusses the construction of the demonstration plant, the mechanical and pre-start-up testing, and the test and evaluation program. A brief description of the process, recent material evaluations and an assessment of application of citrate technology to other industrial SO₂ emitting sources are presented.

CITRATE PROCESS DEMONSTRATION PLANT — CONSTRUCTION AND TESTING —

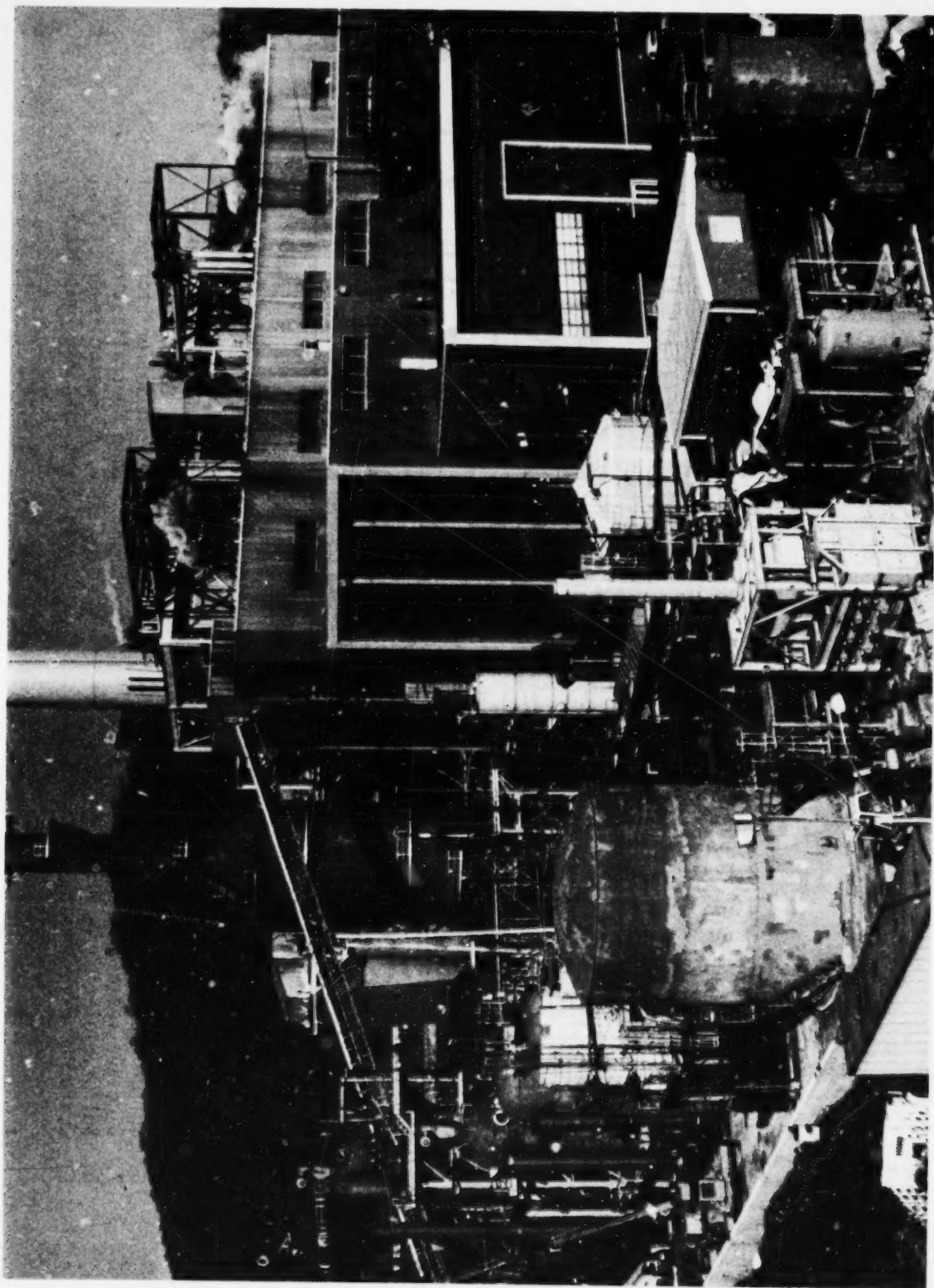
INTRODUCTION/PERSPECTIVE:

Construction and testing of the Citrate Process Demonstration Plant as shown in Figure I, represents an interim milestone in the successful development and demonstration of the Citrate Flue Gas Desulfurization (FGD) Process. The Citrate Process was developed by the U.S. Department of the Interior's Bureau of Mines to meet the goal of minimizing the undesirable environmental impact of industrial plants emitting SO₂ bearing gas. For almost a dozen years, the Federal Bureau of Mines researched, pilot tested and evaluated this technique of aqueous scrubbing with hydrogen sulfide regeneration. Successful pilot testing of the Citrate Process by both government and industry at both metallurgical and coal-fired industrial applications confirmed the technical merits of the Citrate Process over a wide range of differing flue gas characteristics.

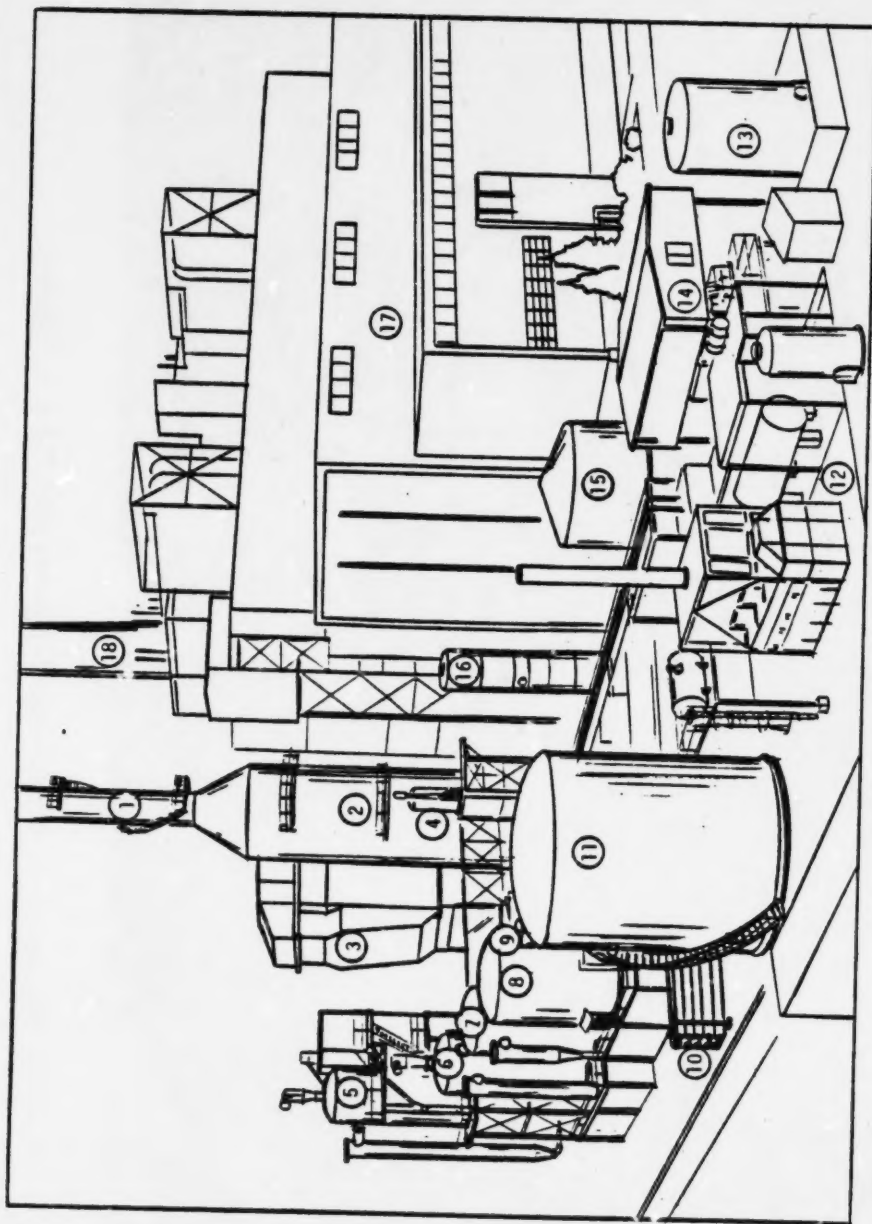
Consistent with the national objectives for energy independence and to derive "Clean Power from Coal", (1)¹ it was decided by a joint government-industry team to demonstrate the Citrate Process at a 60 MWe coal-fired electric power generating plant owned and operated by the St. Joe Zinc Company (St. Joe) of Monaca, Pennsylvania. A cost sharing cooperative agreement between the Bureau of Mines, the U.S. Environmental Protection Agency (EPA) and St. Joe provided the necessary impetus for the engineering, design, construction, and one-year demonstration testing of the Citrate Process at a commercial scale power plant installation. The engineering, construction management, construction and operational testing responsibility are being provided to the project principals by the Morrison-Knudsen Company, Inc., of Boise, Idaho.

The primary objective of the Citrate Process Demonstration Plant project is to demonstrate that the Citrate Process can reliably and efficiently remove sulfur oxides from the flue gases at a commercial scale power plant installation. To achieve this objective, the Citrate Process was installed to treat a nominal 156,000 SCFM (180,000 SCFM, maximum) of SO₂ bearing flue gases from St. Joe's G. F. Weaton power station, as shown in Figure II. The existing flue ducting configuration allows for transfer of additional untreated flue gases from an adjoining twin boiler to the Citrate FGD unit without physical modification of gas ducting. The St. Joe power plant operates essentially base loaded and provides electrical energy to St. Joe's adjoining zinc smelter and to the local power grid. The power plant's main steam flow is 450,000 lb/hr at 1000° F, 1850 psig and is provided with integral reheat. The station is operated on a continuous basis and load changes are typical of an electric utility. The overall station heat rate is approximately 10,200 Btu/kWh.

¹ Underlined numbers in parentheses refer to the list of references at the end of this report.



Overview Photograph of Constructed Demonstration Plant - Figure I



LEGEND

1. Flue Gas Monitoring Platforms/Stack
2. SO₂ Absorber
3. Electrostatic Precipitator
4. Sodium Sulfate Crystallizer
5. Sulfur Precipitation Reactor No. 1
6. Sulfur Precipitation Reactor No. 2
7. Digestor Vessel
8. Sulfur Flotation Tank
9. Sulfur Slurry Tank (Not Visible)

10. Sulfur Melter
11. Citrate Solution Storage Tank
12. Hydrogen Sulfide Generator
13. Fuel Oil Storage Tank
14. CO/CH₄ Compressor House
15. Molten Sulfur Storage Tank
16. Lime Neutralization Tank
17. G. F. Weaton Power Station
18. Existing 275 Foot Stack

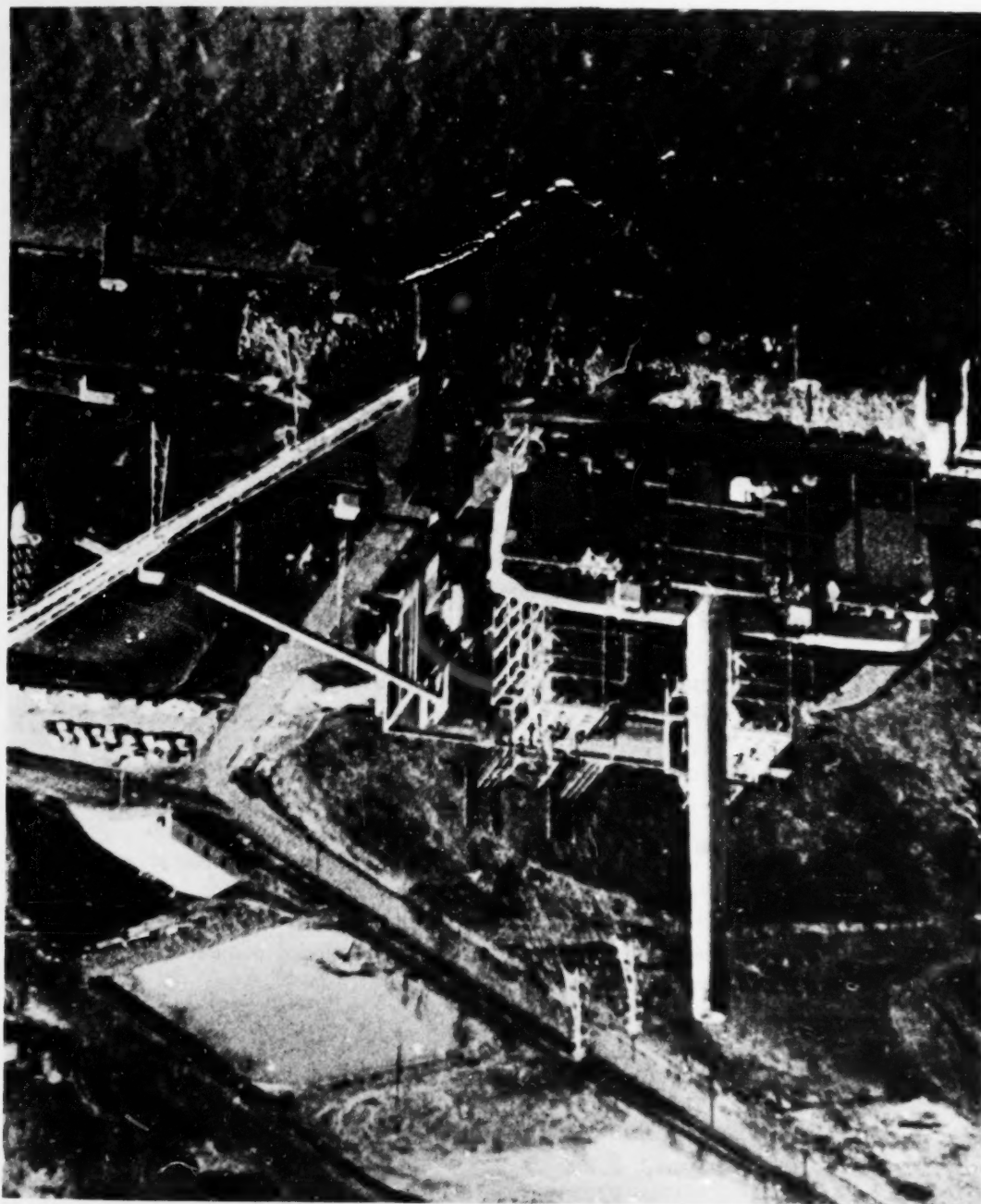


Photo of G. F. Weaton Power Station

Figure II

PROGRESS:

A network diagram of the engineering, procurement and construction progress is shown in Figure III. This figure is a simplified critical path analysis and approximate representation of the various engineering and construction activities and work tasks that comprise the effort to design, build, and test the Citrate Process Demonstration Plant.

A summary of activity milestones associated with the project's four separate phases follows:

Phase I — Preliminary engineering and design development necessary to establish a definite construction cost estimate was completed in November, 1976.

Phase II — Final Engineering, detailed design, and equipment procurement of most major equipment and engineered items was completed by March, 1978. Major equipment including columns, vessels, tanks, reactors, pumps and exchangers was received on site by September 1978. Installation of process and utility piping, instrumentation, electrical work, insulation, and flake glass lining of process vessels was completed in early 1979. Preliminary mechanical testing of the demonstration plant and completion of most construction activities was achieved by March, 1979.

Phase III — Plant testing was initiated in March, 1979, and is expected to be completed by June, 1979.

Phase IV — A one year demonstration testing and performance evaluation program will be conducted by the Radian Corporation (Radian) of Austin, Texas, which has been retained by the Bureau of Mines as an independent testing and evaluation contractor. The one year demonstration test is scheduled to commence upon completion of performance testing and plant acceptance, after which time St. Joe fully expects to continue operation of the Demonstration Plant to achieve continued compliance with applicable environmental regulations.

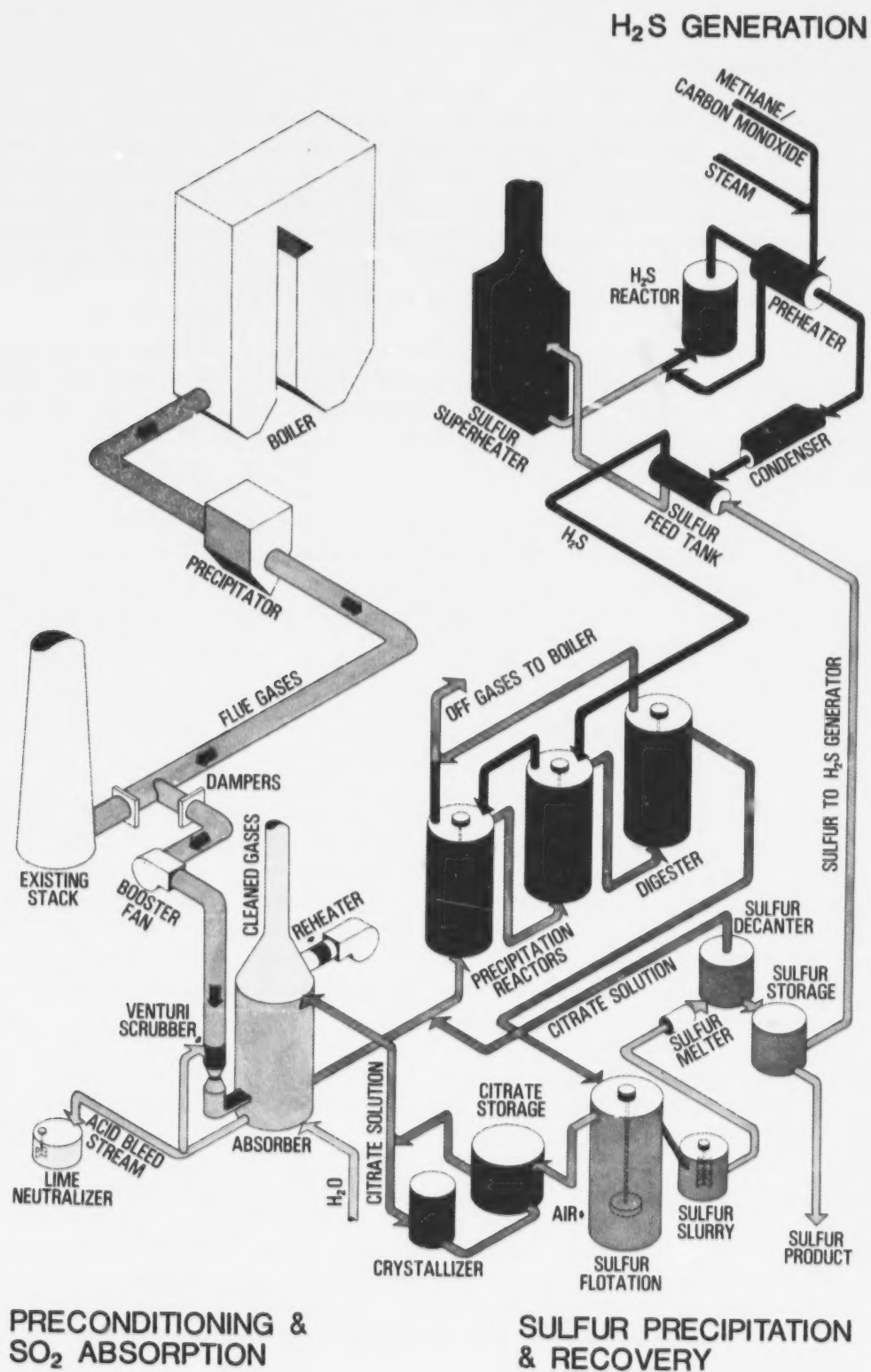
THE CITRATE PROCESS:

Citric acid has been demonstrated as an effective buffer for the aqueous absorption of sulfur dioxide. The absorption/regeneration system using this organic acid is known as the Citrate Process. Citrate chemistry as well as a detailed description of the process has been previously reported and is cited in the references. (2-5)

The Citrate Process for sulfur dioxide emission control comprises five basic steps:

1. **Preconditioning the entering flue gas.** Depending upon the application, the entering flue gases may require reductions in the levels of chlorides, sulfuric acid mist, and particulates, as well as quenching to adiabatic saturation temperatures.
2. **Absorption of SO₂ in an aqueous buffered solution of organic acid.** Conditioned flue gas enters the SO₂ absorber where it flows counter-current to descending citrate solution.
3. **Reaction of the SO₂ loaded solution with H₂S in a closed vessel to precipitate elemental sulfur.** Sulfur dioxide absorbed in the citrate solution is reduced to elemental sulfur by reaction with H₂S, thus, regenerating the absorbent solution.
4. **Separation of the sulfur from regenerated solution.** The elemental sulfur precipitate is concentrated by air flotation into a sulfur slurry which is separated from the regenerated solution. The sulfur slurry is heated to form liquid sulfur to enable decantation from the retained citrate solution.
5. **H₂S generation.** The H₂S required for use in regeneration is either obtained as a by-product of petroleum refining or produced on-site by reaction of recovered sulfur product with a reducing gas and steam.

The basic processing unit operations associated with the Citrate Process are diagrammatically presented in Figure IV.



Citrate Process Flow Diagram - Figure IV

DEMONSTRATION FACILITY:

The Citrate Process Demonstration Plant is designed as a single process train consisting of flue gas pre-treatment, SO₂ absorption, sulfur precipitation, sodium sulfate removal, and H₂S generation. Mechanical and process equipment within the demonstration plant is capable of being scaled upward so that once demonstrated in the plant, the process can be applied with confidence to larger coal-burning utility, refinery, or smelter installations. Typically, equipment selected for the demonstration plant has been a successfully tested in similar commercial applications. Equipment sizing is representative of components for process trains associated with a 500 MWe power plant installation.

As a demonstration plant, a conservative design approach was followed emphasizing the use of specific knowledge acquired through operation of two previous Bureau of Mines' citrate pilot plants in anticipating and preventing problems. Certain optional process equipment and technical features were provided to enhance process flexibility, optimization, and to avoid potential problem areas. Design features included in the constructed plant which emphasize this philosophy follows:

- A low energy venturi scrubber was provided upstream of the absorber column to remove a substantial percentage of chlorides, SO₃ and NO₂ which could interfere with operation of the process. The venturi scrubber also functions to cool incoming gases to achieve efficient absorption.
- To prevent corrosion problems, the ductwork and the flue gas fan were located upstream of the venturi scrubber thus being exposed to only hot, dry gas above the acid dew point.
- The demonstration plant can be tested to ultimate capacity. The flue ducting configuration allows for transfer of additional untreated flue gases from the adjoining twin boiler to the Citrate FGD unit without physical modification of gas ducting. Approximately 18 percent excess fan capacity is provided that will allow for gas treatment during overload conditions. Similarly, the SO₂ absorber, H₂S generator, and other process limiting unit operations are designed for overload conditions.
- Both indirect hot air reheat with steam and an external combustion oil-fired reheater are provided to enhance system reliability and to permit evaluation of these alternative reheat methods.
- Rich absorbent solution is pumped from the absorber to the highest of two stirred precipitation reactors and the solution then flows through the other reactor, digester, and flotation tanks by gravity. The reactors feature a turbine and gas sparger design which will not be subject to plugging. Agitation is maximized to provide good reaction with H₂S gas; the degree of reactor agitation can be adjusted by gear modification.

- Experience has shown that excess H_2S tended to cause difficulty in the flotation of sulfur. Therefore, the stirred digester tank was provided with excess capacity to allow time for mixing and reaction of any excess H_2S with a small flow of rich citrate solution which by-passes the reactors. This gives better flotation and makes subsequent handling of the regenerated solution safer.
- Separation of sulfur from the absorbent solution and subsequent melting was successfully demonstrated during pilot plant testing. The method, based on air flotation of sulfur, is similar to existing technology used in the Stretford process. All lines which contain molten sulfur are steam jacketed and insulated to prevent plugging.
- The oxidation of absorbed SO_2 is expected to be less than 2%. Excess sulfate in the system will be removed as Glauber's Salt in a vacuum crystallizer, which will keep the sulfate level sufficiently low so that crystallization will not occur in the process lines.
- The H_2S generator was designed to minimize carry-over of sulfur vapor into the H_2S product lines. An alternating steam heated/water cooled heat exchanger was provided to condense out sulfur carry-over. In the event of the sulfur carry-over the H_2S gas lines are steam jacketed so that any sulfur will be transported to the precipitation reactors for recovery.
- Due to the corrosive nature of the absorbent solution and the possibility of chlorides being present, construction materials for the demonstration plant were carefully chosen. Hastelloy C-276 or Inconel 625 were used for agitators, pumps and load bearing clips. Piping is FRP, polypropylene lined steel or rubber lined steel. Tanks are lined with rubber or vinylester flake glass. Some smaller tanks are FRP.
- The two sulfur precipitation reactors are provided with by-pass capability to enable continuous process operation with only one reactor in operation; this feature will permit optimization of H_2S utilization.
- The H_2S generator is designed to operate on carbon monoxide as an alternate reducing gas. However, operation with a CO feedstock will require the installation of a gas transport pipeline from the adjoining smelter, where CO is an available by-product from the electrothermic reduction of zinc.

CONSTRUCTION:

The decision to proceed with the detailed design and construction of the Demonstration Plant was made in March, 1977. Materials and special components such as Hastelloy clad vessels and Inconel-625 agitators were promptly ordered and the fabrication of special process equipment and major vessels was started.

Shop fabricated vessels including the SO₂ absorber, sulfur precipitation reactors, digester, sulfur slurry tank, and citrate storage tank, and H₂S generator reactors and vessels were received at the project site in June, 1978.

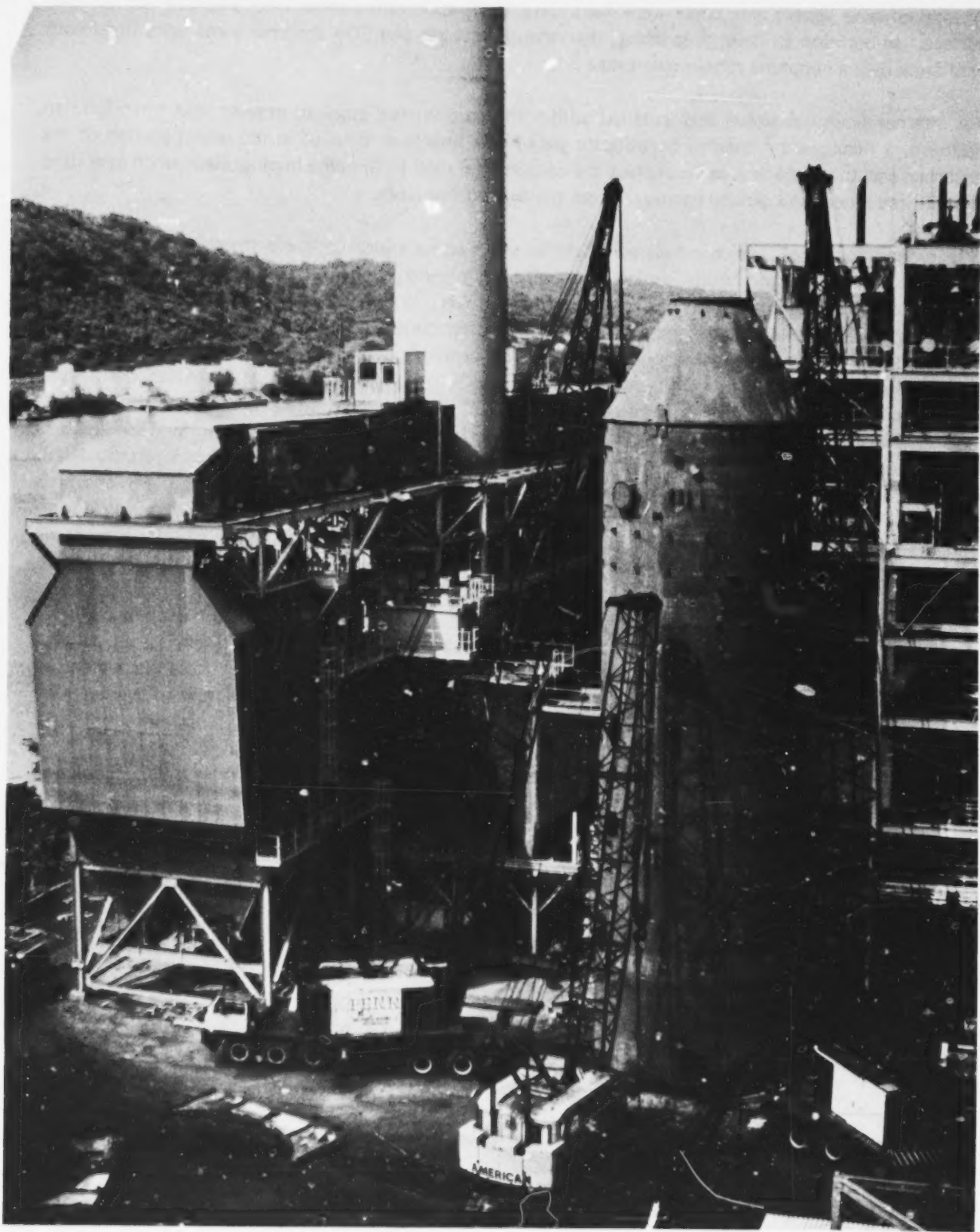
26-foot diameter by 98-foot long SO₂ absorber and the 102-foot long by 10-foot diameter top mounted stack were both shop-fabricated and delivered to the site as single units. However, the absorber internals including the chevron mist eliminator, liquid distributor channels, absorber packing, and reheater were all field installed.

Field erection of the SO₂ absorber, stack assembly, and major vessels required considerable planning and site preparation. The vessels were shipped by barge approximately 1200 miles, off-loaded, and transported two miles on special vehicles before being field erected and set into position on foundations.

The absorber section required significant rigging to enable field erection of this 150-ton vessel with a single lift. Erection of the absorber vessel is shown in the accompanying photograph, Figure V.

Erection of the SO₂ absorber was the critical path in the construction schedule because numerous related work activities such as installation of flue gas fan, ducting, and scrubber recycle pumps were restrained until the absorber was erected. The project critical path was further complicated by the late delivery of the absorber and other major vessels, fabricated lined pipe, and by limited availability of skilled craftsmen (pipefitters). Although the major process vessels and columns were promised for delivery in December 1977, the actual equipment did not arrive on site until mid-June, 1978. Late material deliveries were also experienced with the factors coupled with a high demand for skilled craftsmen in the local labor area resulted in significant loss of potential production and contributed directly to delaying the overall completion of the project by approximately six months.

All wetted process surfaces inside the steel vessels were lined with a flake glass vinylester to protect the carbon steel from corrosion attack by an allowable two weight percent chlorides contained in the citrate solution. To assure that the protective lining would withstand the corrosive environment for the design plant life of approximately 20 years, the metal surfaces to be coated were sandblasted to white metal and immediately primed; a layer of vinylester glass, approximately 35 mils, was then troweled and rolled, and a second layer of vinylester glass, approximately 30 mils, was spray applied. Field quality control consisted of continuous visual inspection while lining installation was in progress, random magnetic thickness measurement, and a 100 volt/mil spark test for the entire lined surface.



Erection of Absorber Vessel - Figure V

A total of nine vessels and tanks were flake glass lined representing some 17,000 square feet of lined surface. In addition to flake glass lining, the venturi scrubber and SO₂ absorber sump were lined with acid brick over a neoprene rubber membrane liner.

An internal fiberglass sleeve was installed within the carbon steel stack to prevent acid corrosion. In addition, a Pennguard² foamed borosilicate glass block liner was installed in the reheat section of the absorber and tower stack area to protect the carbon steel shell from flame impingement when operating the oil-fired reheat and general corrosion from the treated flue gases.

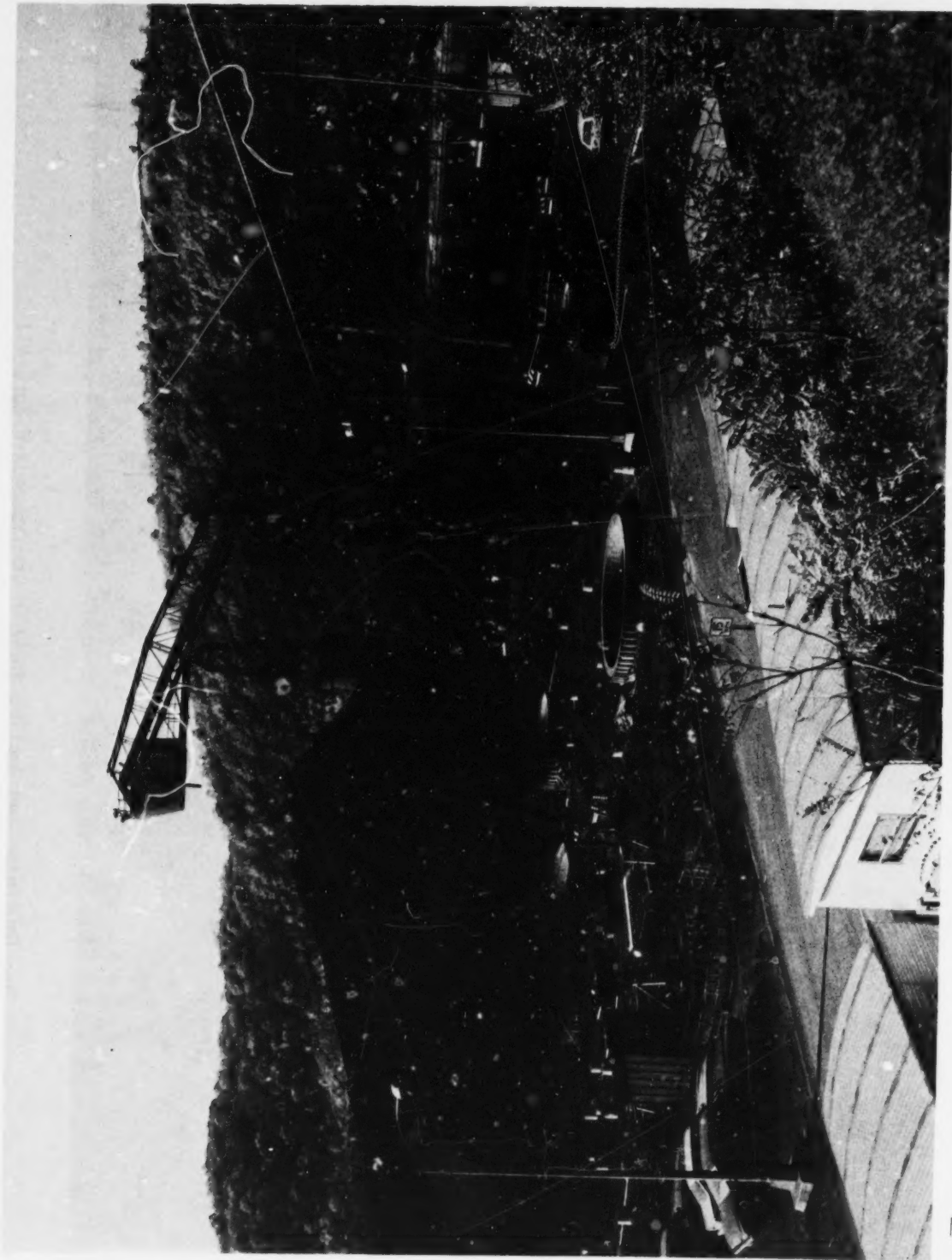
Site preparation work, which consisted of demolishing some minor concrete structures, relocation of a power plant sump and utility piping, and identifying underground interference within the demonstration project's battery limits, was completed by October, 1977, at which time installation of underground and yard piping was started, and shortly thereafter construction of equipment foundations was begun. Sixty-nine pour-in-place concrete piles were installed, two process sumps and a total of 100 equipment foundations were built. Approximately 1500 cubic yards of concrete and 126,000 lbs. of structural rebar were used. About three and one-half months were required to complete the concrete and civil work activities which are shown in Figure VI. Installation of the structural steel supports for the sulfur precipitation reactors, digester, and air flotation vessel was completed in February 1978. The process and utility pipe-rack, a pipe bridge, structure for the sulfate crystallizer system, and H₂S generator, and ducting support structure, as shown in Figure VII, were installed intermittently due to scheduling restraints. Total time for the structural activities was approximately seven months; in all some 550,000 lbs of structural steel, anchor bolts, and miscellaneous steel were consumed by the project.

Approximately 15,000 feet of utility and underground piping were installed at the project site; additionally, some 8,500 feet of process piping, mostly carbon steel polypropylene lined pipe and steam jacketed pipe, were installed. All process and utility piping 2-inch diameter and less were field fabricated and all larger diameter piping, lined pipe, and related fittings, hangers, and supports was completed over a period of eight months.

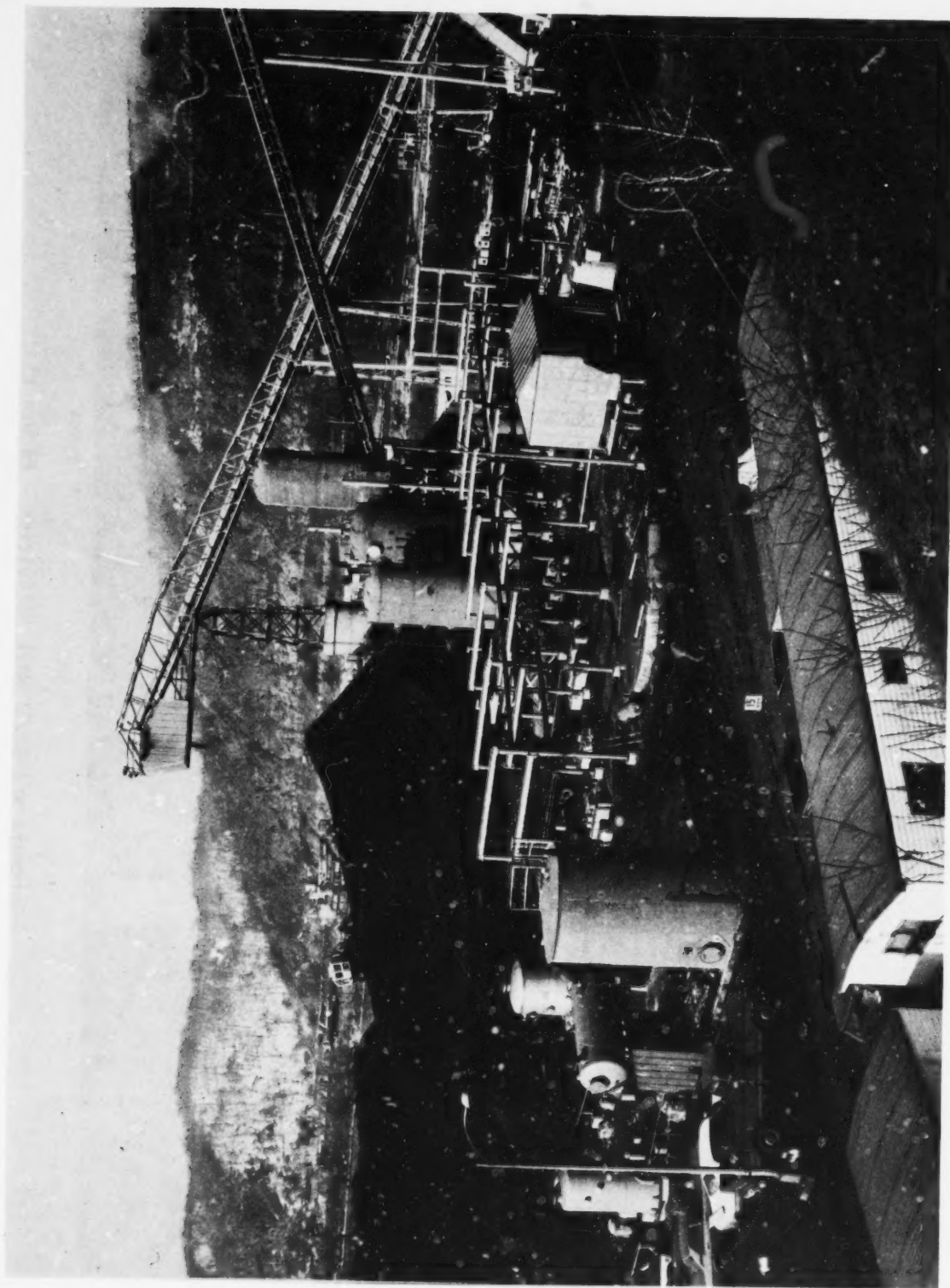
Utility, steam, flue gas, and process off-gas tie-ins to the G. F. Weaton power station were scheduled to coincide with the power plant annual shutdown.

Electrical work consisted of installation of 22,000 feet of conduit, 1,300 feet of cable tray, 112,000 feet of power cable, equipment grounding, area lighting and power service, and the installation of five 2300 volt motors and sixty-four 600-volt or smaller motors. In addition, switchgears, power distributor centers, transformers, and an all-electronic instrumentation package consisting of some 200 sensing elements, transmitters, and recorders were installed. The installation of the electric work and instrumentation occurred over a five month period.

² Reference to specific trade names is made for identification only and does not imply endorsement by the Bureau of Mines.



Concrete and Civil Work Activities - Figure VI



H₂S Generator and Ducting Support Structure - Figure VII

All field construction was performed by journeymen and apprentice craft labor. Some twelve different labor skills were used; these included pipefitters, iron workers, carpenters, masons, millrights, electricians, boilermakers, insulators, painters, and laborers. The peak project manloading for all craft and subcontractors was 102 men. A total of approximately 150,000 manhours of craft labor was expended in the construction of the demonstration plant.

Concurrent with the completion of construction of different process systems, mechanical testing of each component, vessel and related pipe network was made to assure completeness and operational integrity.

MECHANICAL TESTING:

Mechanical testing of the various mechanical and process components, piping systems, and auxiliaries was initiated during February, 1979.

Mechanical testing implemented to avoid plant start-up problems, consisted of visual inspection of all equipment and piping systems to assure completeness, spark testing of all lined vessels, and hydrostatic leak testing of pressure vessels, solution tanks, and pipe networks. Hydrostatic testing of jacketed, lined and specialty piping and stress relieved vessels was conducted in the fabrication shop prior to release for field installation. Once installed, entire systems were then systematically pneumatically pressure tested and/or hydraulically tested, depending upon service. In the case of the refractory lined H₂S generator vessels and piping, a halogen leak test was used to avoid wetting the refractory surfaces.

All electric motors, switchgear, and distribution apparatus were functionally tested to assure proper motor rotation, and to determine workability of motor controls and safety interlocks. Similarly, all electronic instrumentation was field calibrated and tested over the intended operating range. Control valves, regulators, and safety relief valves were all "stroke" tested and adjusted to performance specifications.

Process lines were water washed to remove debris and foreign matter; steam and condensate lines were degreased and steamed-out. Unlined carbon steel tanks for caustic and molten sulfur storage were cleaned of loose mill scale and corrosion products.

PERFORMANCE TEST PROGRAM:

Testing of the demonstration plant will establish Citrate FGD technology in mechanical and process equipment of sufficient size so that scale up to 500 and 1000 MWe power plants is feasible, and that such applications can be built and the performance, capital, and operational costs predicted with confidence. To insure the objectivity of the results, the Bureau of Mines solicited proposals for independent testing and evaluation of the Demonstration Plant. Radian Corporation of Austin, Texas, was the successful bidder and was selected to perform the evaluation.

Testing will be done by operating with the demonstration plant in a variety of operating modes of varying loads, while burning various coal fuel mixtures with sulfur contents between 2.5 to 4.5 weight percent. The test program consists of baseline testing, acceptance testing, and a one-year demonstration program.

- **Baseline Test** — Baseline performance testing of Unit No. 1 of the G. F. Weaton power station with selected coal fuel was conducted to characterize the flue gas in the untreated mode prior to start-up of the demonstration plant. The baseline test was completed during November, 1978. These tests characterized the boiler, particularly with respect to flue gas emissions, sulfur balance, and combustion efficiency. Results of the baseline test serve as a basis for evaluating the effectiveness of the Citrate Process and for determining what effect, if any, retrofit of the demonstration plant has on boiler operation. Preliminary results indicating the flue gas emissions and the coal sulfur content and heating value are presented in Table I.

Information gathered during baseline testing includes:

- Definition of the characteristic operating boundaries of the steam generating facility. Historical operating data on normal load fluctuations, excess air requirements and fuel analysis were studied, along with the boiler reliability of the George F. Weaton Power Station.
- Determinations of the relationship between generating facility control settings and operating conditions of the boiler and the resulting emissions.
- Collection of quantitative data on pollutants for the purpose of establishing realistic emission parameters for varied operating conditions.
- Determination of realistic baseline operating parameters as required for the accurate and economical operation of the Citrate FGD Process.

**TABLE 1. Preliminary Results of Baseline Testing
Unit No. 1 — G. F. Weston Power Station**

GAS ANALYSIS (to stack)		COAL ANALYSIS (dry basis)	
SO ₂	1990 PPM	Ash	13.1%
SO ₃	8 PPM	Sulfur	3.04%
NO _x	350 PPM		
Cl	70 PPM		
Particulates	0.024 Gr/DSCF		

- **Acceptance Test** — For the purpose of fulfilling the contractual agreements relative to system performance and operation, a ten consecutive day test of the Citrate Process Demonstration Plant will be made. During the acceptance test, the G. F. Weston station will be fired with coal containing 3.0% percent sulfur. Performance and operating parameters to be measured during the acceptance test include:
 - Sulfur content of coal fuel.
 - Average boiler load.
 - SO₂ removal efficiency, percent.
 - Sulfur product output, STPD.
 - SO₂ in outlet flue gas, PPM and lbs/MM BTU input.
 - Sulfur assay.
 - Electrical power requirements.
 - Steam requirements.
 - Particulate in absorber inlet flue gas, Gr/SCF.
 - Particulate in absorber outlet flue gas, lbs/MM BTU input.
 - Process feedstock requirements.
 - Sulfate concentration in lean absorber liquor.
 - Reheat capability

- **One-Year Demonstration Test** — Following the completion of plant performance testing and final acceptance, a one year demonstration test will be initiated. The objectives of the demonstration test are:
 - To characterize completely the Citrate FGD system with respect to the various system operating parameters.
 - To determine the system's optimum operating conditions.
 - To establish the long-term system reliability.
 - To assess the environmental impact of the Citrate FGD system.
 - To define the technical and economic feasibility of the Citrate FGD system.
 - To document the results of the test program such that comparisons of the Citrate system with other flue gas desulfurization systems can be performed.

To accomplish these objectives, the one year demonstration program is designed to determine what adjustments or process refinements are required to minimize the absorber liquid/gas ratio; system pressure drop; feedstock consumption; power consumption; reducing gas consumption; and process capital and operating costs; and to maximize sulfur oxide removal; particulate removal; system reliability; and system availability.

The one year demonstration test will be divided into three periods of approximately four months each. Each period will consist of about two months of optimization testing followed by about two months of steady-state testing.

During optimization testing, parameters will be changed frequently and the response of the system to these changes will be noted. The removal of SO_2 will not be constrained during these tests. The more promising operating conditions identified will be examined further in the steady-state runs.

The steady-state runs will be divided into two one-month test periods, each of which will involve operation at different steady-state conditions. During these tests the removal of SO_2 will be set at 90 percent and other operating conditions will be varied. Parameters which can only be measured during longer term testing (e.g., sulfate formation) will be investigated during steady-state testing.

A summary of the operational and performance tests to be conducted during the one year program follows:

- **First Optimization Test Series** — During the first optimization test series, emphasis will be placed on characterizing the absorber for SO_2 removal and the prescrubber (venturi) for particulate removal. A preliminary examination of the regeneration reactor section will also be carried out. Because the SO_2 absorber and the reactor section are closely coupled through the absorber feed rate and the amount of SO_2 in the citrate liquor leaving the absorber, a significant number of reactor section tests will occur naturally. Items of lesser importance that will also be evaluated include the magnitude of the rich citrate bypass, flotation air requirements, and turndown.
- **First Steady-State Test Series** — Two long-term tests will be conducted during the first steady-state test series. Selection of the operating conditions will be based on the results of the first optimization test series. However, the variables to be emphasized are the absorber L/G and the $\text{H}_2\text{S}/\text{SO}_2$ ratios.
- **Second Optimization Test Series** — The second optimization test series will stress the regeneration reaction section, the ability of the system to follow load, the importance of temperature in the absorption and regeneration steps, and the effect of citrate concentration.
- **Second Steady-State Series** — This test campaign is very similar to the first steady-state test series. Again, the variables emphasized will be the absorber L/G and the $\text{H}_2\text{S}/\text{SO}_2$ ratios. However, the tests will be run with a reduced citrate concentration and the effect of reheat will be studied. Also, the effect of operating the absorber and regeneration reactors at an off-design temperature for longer time periods may be examined.

- **Third Optimization Test Series** — The third set of optimization runs will examine variations in reactor configuration, power plant load following, operation on higher sulfur coal fuel, and the effect of kerosene addition to the flotation tank.
- **Third Steady-State Test Series** — Conditions for the final steady-state run will be selected based on the cumulative experience gained testing the unit. The SO₂ removal will be fixed at 90 percent during this period and the citrate plant will be required to follow the load on the boiler. This final steady-state run will characterize the citrate process at the optimum conditions that can be achieved at the demonstration plant. Chemical and utility consumption, unit availability and reliability, and sulfate formation rate at a given SO₂ removal rate will be studied during this period.
- **Special Studies** — Some aspects of the demonstration testing of the citrate process require special study. These aspects are discussed separately because they do not directly influence the absorption or regeneration steps and, therefore, are not readily examined with the approach used for routine demonstration testing. The areas of special concern during the demonstration tests are sulfate formation, corrosion, and reheat requirements.
 - **Sulfate Formation** — One of the advantages of the Citrate Process is the suppression of sulfate formation. The total sulfate being introduced into the Citrate Process will be monitored. The sources of sulfate include:
 - Sulfuric acid mist in the inlet flue gas.
 - Air oxidation of sulfite during absorption.
 - Disproportionation during regeneration.
 - Air oxidation during flotation.
 - Formation in the melting/decanting stage.
 - **Corrosion** — The materials used in the FGD system are important factors in determining both the reliability of operation and the capital costs. In the demonstration plant, emphasis is on reliability, and therefore, higher quality materials than may be necessary were specified. Operation of the test facility will provide an opportunity to test and evaluate other materials which may ultimately lower system costs or provide a backup in case of failure of present materials. The corrosion topic is further discussed in the material evaluation section of this paper.

- **Reheat Requirements** — The effectiveness of flue gas reheat in the Citrate system by air injection from a steam/air heat exchanger will be assessed. An oil-fired burner type reheat system was also installed to allow a technical and cost-effective comparison of the indirect steam reheat versus oil reheat.

MATERIALS EVALUATION:

Corrosion testing was begun at the Albany Metallurgy Research Center of the Bureau of Mines following observation of corrosion within the unit operations associated with the Citrate Process pilot plant. Apparatus was assembled in the laboratory to simulate the following environments of the demonstration plant:

- Sulfur vapor at 1,150° F to 1,450° and 1 atmosphere pressure. An alternate atmosphere in this category is sulfur vapor plus steam in the same temperature interval.
- Gaseous atmosphere $H_2S-CO_2-H_2O$ at 200° F to 1,150° F and 1 atmosphere pressure.
- Rich citrate solution in the absorber containing 5 to 15 g/l SO_2 at 120° F and ambient pressure.
- Lean citrate solution in the sulfur melter and decanter containing no SO_2 at 260° F and 4 atmospheres pressure.

A summary follows of test results obtained to-date. The investigation is not complete except for evaluating refractory materials in sulfur vapor and sulfur vapor containing steam. All other results must be considered tentative pending replication in future tests. Final reports will be issued as Bureau of Mines Report of Investigations.

Laboratory Tests in Sulfur Vapor and Sulfur Vapor Containing Steam

Three castable refractories were evaluated for application in the H_2S generator. Cylindrical specimens (3/4 in. dia. x 1-1/4 in. long), after exposure for 30 days at 1,450° F to sulfur vapor, steam and sulfur vapor containing approximately 50 percent steam, were tested for compressive strength and compared with similar specimens heated in air under the same conditions of time and temperature. The data, as shown in Table 2, indicates significant loss of strength for High Strength Brik Cast G in sulfur plus steam. A.P. Green Co. KS-4 and Gunning Refractories Co. HYDRA-SEH4 were satisfactory under all test conditions.

TABLE 2. Compressive Strength (psi) of Castable Refractories After 30 Days at 1450° F

Material	Test Atmosphere			Steam
	Air	Sulfur	Sulfur 50 pct. Steam	
A.P. Green Co. KS-4	2250	2100	5500	3450
General Refractories Co. High Strength Brik Cast G	3650	5000	550	4450
Gunning Refractories Co. HYDRA-SEH4	3850	3250	2200	2600

Most metals were severely attacked by sulfur vapor at high temperature, as shown by the preliminary test results listed in Table 3. Chromium is the principal element deterring sulfidation of iron-base alloys, and nickel provides additional resistance at 1,150° and 1,250° F. The austenitic stainless steels were attacked more than ferritic alloys at 1,350° F, probably because of the tendency for nickel to lower the melting temperature of sulfide scales.

Alonizing, a diffusional process which enriches a metal surface in aluminum, was very effective for decreasing sulfidation of iron-base alloys.

Titanium was found to be very resistant to sulfur vapor in a single test at 1,250° F.

**TABLE 3. Corrosion of alloys in 1 atm. Sulfur Vapor
mils per year (mpy)**

Alloy	Temperature, °F		
	1150	1250	1350
Carbon Steel	>1400 ⁴	—	—
Carbon Steel, Alonized	—	4	19
CRM-4 (Fe-6Al) ³	—	>250 ⁴	—
Type 406, Alonized	—	5	—
Type 446	120	140	80
E-Brite 26-1	—	115	80
Type 316	43	110	160
Type 316, Alonized	—	9	12
Type 310	43	115	205
Type 310, Alonized	—	7	27
Hastelloy C-276	—	42	—
Titanium,	—	27	—

Laboratory Tests in H₂S — 16 pct. CO₂ — 16 pct. H₂O at 200° to 1,150° F

Remarkably similar corrosion rates for iron and nickel-base alloys containing greater than 12 percent chromium were measured in the subject atmosphere with and without intermittent exposure to steam and thermal cycling to room temperature. Corrosion rates for Incoloy 800; Inconel 601; types 304, 316L, 317L, 310, and 446 stainless steels; E Brite 26-1; Uniloy 18-2; and MP-35N lie within the following ranges at the temperatures indicated; 600° F, < 2 (mpy); 800° F, 5±4 mpy; 950° F, 20±15 mpy; 1,075° F, 50±30 mpy; and 1,150° F, 100±50 mpy.

Aluminum as a binary additive to iron was not beneficial in the test atmosphere, but it reduced sulfidation considerably in combination with chromium. For example, Armco 18SR (18Cr-2Al), type 406 (13Cr-4Al), and Hoskins 815 (26Cr-4 Al) formed adherent sulfide scales and corroded at 37, 10, and 6 mpy, respectively, at 1,150° F.

³ Experimental iron base alloy of Chrysler Corp.

⁴ Specimens were penetrated in the test.

Laboratory Tests in Citrate Solution

Stainless steels and Fe-Cr-Ni base alloys including Incolloys, Inconels and Hastelloys as well as Ti and Zr were exposed for 30 days in static salty citrate solutions containing either 10 g/l SO₂ (rich Citrate) or less than 0.2 g/l SO₂ (lean Citrate). The nominal composition of the test solutions and other conditions are listed in Table 4. Three coupons 1-1/2 in. x 3/4 in x 1/16 in. were exposed in 400 ml of rich citrate solution through which flue gas was passed via a bubble pump to effect stirring. One of the coupons was half immersed, the second fully immersed, and the third was fitted with a nylon bolt, nut, and washer to invite crevice corrosion. Glass vessels containing 500 ml of lean citrate solution and two coupons were pressurized with air. Both coupons were fully immersed, and one was provided with a crevice. General and localized corrosion (crevice corrosion and pitting) were determined by weight loss, by visual examination at low magnification, and by metallography.

TABLE 4. Test Conditions and Nominal Composition of Citrate Solutions

Component	Rich Citrate Solution	Lean Citrate Solution
Citric Acid	96 g/l	96 g/l
NaOH	40	40
Na ₂ S ₂ O ₃	63	63
Na ₂ SO ₄	71	71
NaCl	83	83
SO ₂	5 to 10	<0.2

Atmosphere	Flue Gas Bubbled	
	Through Solution	Pressurized with Air
pH	4.2 to 4.3	4.4 to 4.5
Temperature, °F	120	260
Pressure, atm	1	4
Time, Days	30	30

Unmodified ferritic stainless steels and ferritic stainless steels containing 2 to 4 percent Al or 1 to 2 percent Mo corroded excessively (> 100 mpy) and are not suitable for use in either lean or rich citrate solutions. In general, the austenitic alloys corroded with rates less than one mpy and differed primarily in their resistance to localized corrosion. As shown in Table 5, the resistance of Fe-Cr-Ni base alloys to localized corrosion increased with Mo + W content. In rich citrate solutions at 120° F, alloys containing 6 percent (Mo + W) were resistant. In lean citrate solutions at 260° F, Fe-Cr-Ni base alloys containing less than 6 percent (Mo + W) were still susceptible to localized corrosion, and only alloys containing 9 percent or more (Mo + W) were completely resistant. Ti and Zr were resistant to corrosion in both rich and lean citrate solutions.

**TABLE 5. Resistance of Alloys to Localized Corrosion
(Crevice Corrosion and Pitting) in Citrate Solutions**

Alloy	Mo %	W %	<i>Tendency for Localized Corrosion</i>	
			Rich Citrate Solution, 120° F	Lean Citrate Solution, 260° F
Hastelloy C-276	16	0	None	None
Hastelloy C-4	15	0	None	None
MP-35N	10	0	None	None
Inconel 617	9	0	None	None
Inconel 625	9	0	None	None
Hastelloy G	6.5	0	None	None
Allegheny Ludlum 6X	6.5	0	None	None
RA 333	3	3	None	Slight
N-155	3	3	None	Slight
Titanium	0	0	None	None
Zirconium 705	0	0	None	None
Sandvik 2RK65	4.5	0	Slight	Moderate-Severe
Incoloy 825	3	0	Slight	Moderate-Severe
Type 317L	3-4	0	Slight	Moderate-Severe
Type 316L	2-3	0	Slight	Moderate-Severe
Type 316	2-3	0	Slight	Moderate Severe
Carpenter 20Cb-3	2-3	0	Moderate-Severe	Moderate-Severe
RA 330	0	0	Moderate-Severe	—
Hastelloy B ⁵	28	0	Moderate Severe	—
Inconel 600	0	0	Moderate-Severe	—
Incoloy 800	0	0	Moderate-Severe	—
Type 304	0	0	Moderate-Severe	—

It should be noted that static conditions might encourage localized corrosion and minimize general corrosion, whereas flowing solutions might increase general corrosion and decrease the tendency for localized corrosion. Consequently, the laboratory test results may disagree with the results of in-plant testing.

⁵ Contains only 1 percent Cr compared to > 12 percent for most other alloys.

APPLICATIONS AND FEATURES:

In addition to coal and oil-fired utility and industrial boilers, the Citrate Process can be applied to sulfur dioxide emission discharges from the refinery and metallurgical industries. (6) Engineering and economic designs for Citrate Process installations can reflect the site specific requirements of smelting processes, and the combustion products associated with petroleum and chemical operations.

At petroleum refineries the H_2S derived from sour crude can be used for citrate regeneration. Thus, the Citrate Process permits refiners to improve the economic balance between use of high sulfur fuels for boiler plant operations and the necessary abatement of SO_2 discharges. A Citrate facility installation would also permit integration of other stack gas desulfurization requirements, such as required for process headers, Claus and acid plant tail gas, and fluidized catalytic cracker emissions. Typically, a centralized Citrate regeneration facility would be provided to serve a variety of point sources located throughout the facility.

Sulfur dioxide emissions from steel mill iron ore roasters can be reduced to elemental sulfur with H_2S derived from coke oven gas (COG). More than half the hydrocyanic acid present in COG is stripped in the wet removal of hydrogen sulfide. HCN must be separated from the H_2S prior to its use in citrate regeneration. Various methods are available for separating HCN including aqueous scrubbing and catalytic cracking. Economic design would include the balancing of H_2S derived for Citrate regeneration versus sulfur dioxide formed in the incineration of the COG.

The economic limits of industrial application for the Citrate Process include treatment of flue gas dischargers ranging from five volume percent SO_2 to as low as 500 PPMV. Greater than 90 percent removal of SO_2 can be achieved and cleaned gas can contain less than 25 PPMV SO_2 . The specific features of the process which permit this range of performance and versatility include:

- SO_2 removal efficiencies exceeding 99.0 percent have been documented.
- The Citrate Process has a high capacity for short-term SO_2 overloads and can accommodate rapid load fluctuations and variable SO_2 concentrations.
- Marketing the elemental sulfur product is less costly in storage and transportation than sulfuric acid. The high purity sulfur product is of feedstock quality for many industrial processes.
- Flue gas conditioning and SO_2 absorption are free of scaling and plugging problems; precipitation of sulfur takes place in the regeneration section outside of the absorber.

- Low rate of oxidation to sulfate. Less than 2 percent of the entering SO_2 is converted to sulfate which can be selectively purged by crystallization; the sulfate is recovered as Glauber's salt.
- Low process energy demand in terms of both process steam and electrical power. A utility application requires approximately three percent of station energy output.
- Minimal environmental impact. There is no significant sludge disposal requirements necessitating ponding or voluminous lagoons. Depending upon the character of the entering waste gas, disposal of particulate and acids is required. The citric acid reagent is a non-toxic biodegradable organic which eliminates risk of further environmental deterioration.
- The citrate solution regeneration facility can be physically separated from the gas quench and absorption operation allowing each to operate independently. This feature is more advantageous at space-limited sites or sites where regeneration is more economically performed at a central location other than the emission source.
- The process can use other organic acids such as glycolic acid in lieu of citric acid, should other organics be shown to have improved performance.
- The citric acid reagent serves as a pH buffer and does not directly enter the absorption chemistry.
- A low absorber liquid-to-gas ratio results in reduced pumping and fan head losses.
- Use of the Citrate Process in applications where H_2S is available without on-site generation will recover the sulfur value of the H_2S gas as well as the absorbed SO_2 .

SUMMARY :

Fulfillment of the objectives of the Citrate Process Demonstration Program will provide an attractive alternative for the abatement of SO₂ discharges to industry. Most importantly, the demonstration test will establish credible basis for informed decisions by industrial and utility managers for application of Citrate technology. Within the next year, the system performance will be documented; the reliability and availability of the Citrate system will be established; and the actual costs associated with plant operations, maintenance, and utility/feedstock consumptions will be reported.

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DESIGN AND COMMERCIAL OPERATION OF LIME/LIMESTONE
FGD SLUDGE STABILIZATION SYSTEMS

Ronald J. Bacskai
Lee C. Cleveland

IU Conversion Systems, Inc.
115 Gibraltar Road
Horsham, PA 19044

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ABSTRACT

Installation and operation of flue gas desulfurization (FGD) systems in coal-fired utility boilers are becoming essential to combat air pollution and meet the Clean Air Act SO₂ emission standards as promulgated by government regulatory agencies. All non-regenerative FGD systems produce large volumes of low percent solids sludges, which create a waste disposal problem for the utilities. These sludges, unless properly treated and disposed, can cause water pollution and damage the environment.

The Resource Conservation and Recovery Act will soon mandate standards for SO₂ scrubber sludge and ash waste disposal. It is expected that the FGD wastes will have to be chemically stabilized to minimize environmental degradation and to produce a structural product to minimize land use.

Stabilization of FGD SO₂ sludge and fly ash waste has, therefore, become an important factor in our country's ability to meet environmental standards with corresponding power generation growth.

Although there are many SO₂ scrubbers currently operating with large volumes of sludge being produced annually, there is limited operating experience on full scale dry stabilization systems. That is particularly true of high sulfur eastern coals. This paper discusses the only four commercially operating plants on eastern high sulfur coal. The stabilization process is the IU Conversion Systems For-O-Tec System which involves dewatering of the scrubber sludge followed by mixing with waste coal ash and chemical additives to produce a structurally stable and environmentally compatible material for landfill, land reclamation, roadbase, liners and other useful applications.

A discussion is provided on the process and plant design considerations which must be evaluated to ensure successful plant operations. Problems encountered in designing and initially oper-

ating SO₂ Waste Treatment Systems for waste materials that were not available for evaluation and testing prior to the completion of the multi-million dollar facilities will be reviewed. The design and operation of these systems involves expertise in material handling, chemical processing, geotechnical engineering and regulatory requirements. Actual "hands-on" operating and maintenance experience is reviewed.

Emphasis will be placed on: 1) plant design problems and solutions for them; 2) variances in SO₂ scrubber sludge and coal ash characteristics and their effect on equipment design, plant operation and chemical stabilization; 3) the handling of the waste materials including the dewatering characteristics of sludges and the controlled feeding of fly ash at higher rates than used previously in any industry; and 4) flexibility of choosing disposal sites with the use of the stabilized material produced in a "dry" system.

Five years in this field and the experience derived from the operation of four major waste treatment systems and from design of eleven additional treatment systems has resulted in a substantial knowledge of all facets of such waste treatment systems and the development of more efficient and more economical processing plants to serve industry.

INTRODUCTION

With the uncertainty of natural gas and petroleum imports, and the growing public concern over nuclear power plants, more attention has been focused on the utilization of coal, the most abundant energy resource in the United States, for generation of electric power. With this expansion of coal usage and the increased necessity for environmental controls, it has become necessary for utilities to install and operate a growing number of flue gas desulfurization (FGD) systems for SO₂ removal. It is estimated that nearly 60,000 MW of FGD capacity will be installed by 1980.*

*PEDCO Environmental — February, 1979

FIGURE 1

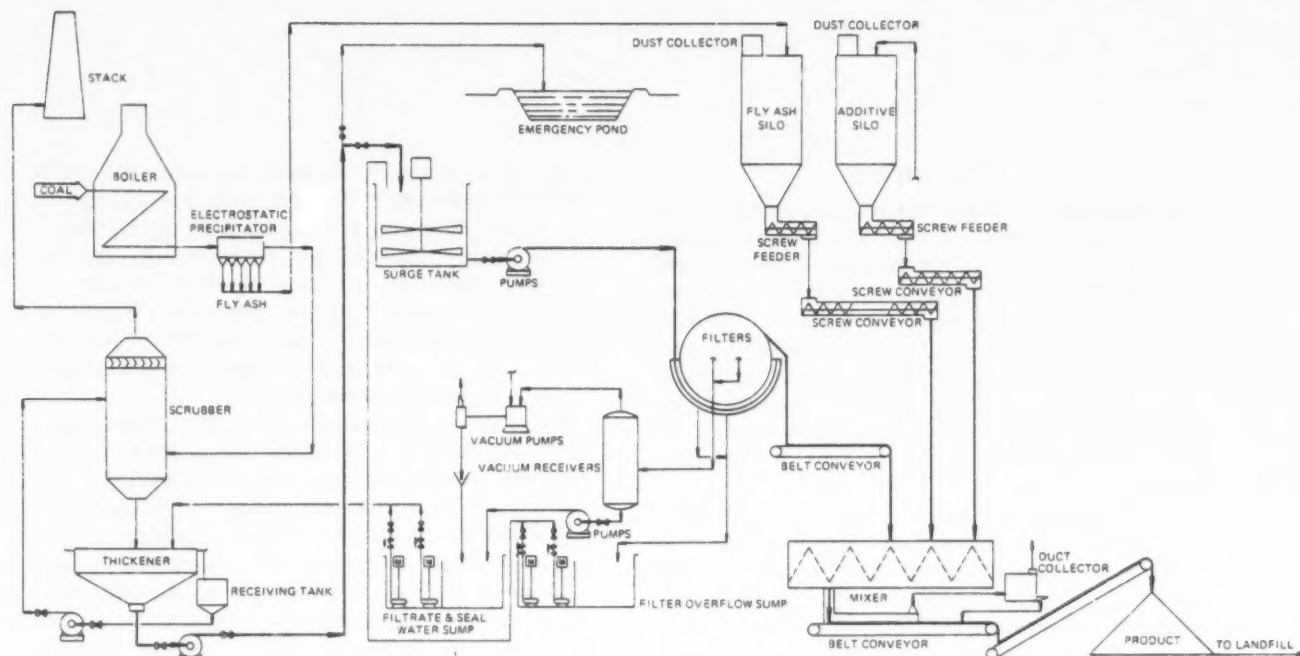


Figure 1 Process flow

Among the various FGD scrubber systems available, wet lime/limestone scrubbing and double alkali (indirect lime/limestone) scrubbing have gained the most industry acceptance. In excess of 90% of all scrubbers installed or committed are of this type. These scrubbing operations produce an enormous volume of low solids content sludge, which must be properly disposed so that ground water and surface water is not polluted by unacceptable concentrations of heavy metals and dissolved solids.

A solution to this massive sludge disposal problem is chemical stabilization of scrubber sludge by the Poz-O-Tec process to prevent significant environmental damage and minimize land disposal requirements. Essentially, the process involves treatment of SO₂ sludge with fly ash and one or more additive(s) to produce, via pozzolanic reactions, a stabilized material, suitable for landfill

disposal. A simplified process flow diagram is shown in Figure 1.

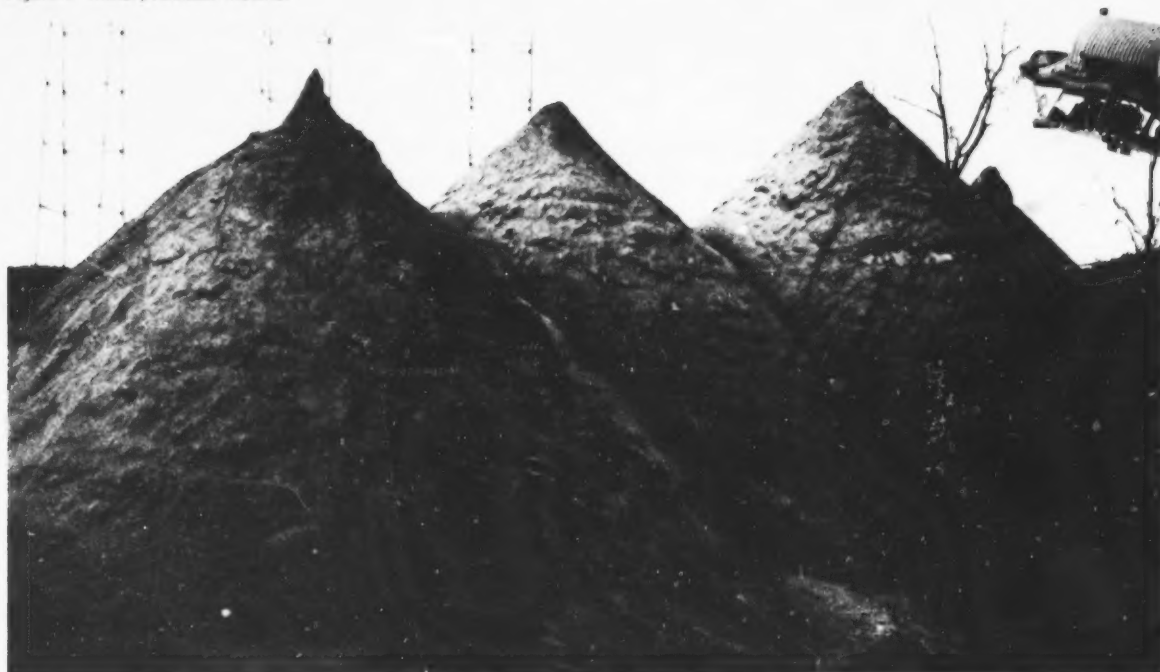
There has been extensive literature written on the Poz-O-Tec process regarding its physical and environmental properties, and this paper will not attempt to address these issues. In summary, the process has been tested on pilot plant applications as early as 1973 at Southern California's Edison Mohave Power Plant; it has been thoroughly evaluated by the EPA at TVA Shawnee with extensive test reports circulated throughout the industry, and it has been tested and evaluated by Louisville Gas & Electric at the Cane Run Plant under an EPA sponsored program. In addition to extensive lab, pilot tests and demonstrations, the Poz-O-Tec system has received large scale commercial acceptance on a wide variety of scrubbers as demonstrated by the Utilities committed to the full scale process applications as shown in Figure 2.

FIGURE 2

IU CONVERSION SYSTEMS, INC.
SO₂ SLUDGE STABILIZATION CONTRACTS

UTILITY	STATION	SCRUBBER	MW	Waste Materials T/Yr.
Big Rivers Electric Corp.	Reid #1 & 2	Lime	480	855,000
Central Illinois Public Service Co.	Newton #1	Double Alkali	575	989,000
Cincinnati Gas & Electric Co.	East Bend #2	Lime	600	1,106,000
Columbus & Southern Ohio Electric	Conesville #5 & 6	Lime	820	995,000
Commonwealth Edison	Powerton #5	Limestone	450	608,000
Duquesne Light Company	Phillips	Lime	400	456,000
Duquesne Light Company	Elrama	Lime	500	629,000
East Kentucky Power Cooperative	Spurlock #2	Lime	500	761,000
Indianapolis Power & Light	Petersburg #3	Limestone	515	620,000
Indianapolis Power & Light	Petersburg #4	Limestone	515	886,000
City of Lakeland	C.D. McIntosh Jr. #3	Limestone	350	470,000
Southwestern Electric Power Company	Henry W. Pirkey	Limestone	720	1,146,000
Texas Municipal Power Agency	Gibbons Creek	Limestone	400	788,000

Figure 3 Initial processed material



The Poz-O-Tec process is a complete waste management system for coal fired power plants. It blends fly ash, bottom ash (if desired), scrubber sludge, lime and other additives. Concentrated streams from the evaporator and cooling tower

sludge can also be incorporated. The stabilized material is a cementitious material, and with proper placement and compaction, exhibits low permeability and superior structural properties. The material in its initial processed form and after final placement is shown in Figure 3 and 4.

Figure 4 Material after final placement



PROCESS CONSIDERATIONS

Sludge disposal has become a major consideration in the design of a power plant and therefore, should not be an afterthought. It should not be left until the rest of the plant has been designed because good sludge disposal is dependent on many aspects of the power plant design including the boilers, coal, type of scrubber, reagent, thickener design, etc.

The stabilization of power plant wastes involves more than just combining the wastes by themselves or with an additive. Each of the waste materials contributes chemically and physically to the process, and variations in those materials must be considered in developing the specific process design.

Fly Ash

Fly ash is utilized in the Poz-O-Tec process for several reasons. It is a waste material which must be disposed of and is usually available at the same source as the sludge waste. It is a fine particle material and provides the alumina and silica which are necessary for the pozzolanic reactions to tie up the sulfur compounds of the sludge.

As the Poz-O-Tec system is a dry stabilization method, the quantity of fly ash will also contribute to the final solids content of the product and effect its handleability. Generally, ash to sludge ratios of 1:1 or higher will result in an immediately placeable material; those below that ratio will usually necessitate interim stockpiling prior to final placement.

Particle size of the ash also contributes to the process chemistry. The finer the particles, usually resulting from modern precipitators, the more reactive the ash. Fly ash from both mechanical collectors and electrostatic precipitators will have a large range of particle size which may reduce its reactivity. Cyclone boilers generate large quantities of bottom ash rather than fly ash. That ash is reactive but is of large particle size and usually collected wet. The ash can be ground to provide adequate particle size for reactivity, but the cost of grinding and the additional moisture make stabilization more expensive and more difficult.

Additives used in the precipitator to improve fly ash collection can cause chemical and safety problems in operating stabilization systems. Ammonia-based additives, for example, can cause a significant reduction in strength characteristics of land-filled material. Further, ammonia gas is released when the ash is in contact with an alkaline material, such as lime. The quantities of ammonia released are considerably higher than permitted by OSHA and such processing facilities would require

extensive air cleaning equipment. Even then, the ammonia odor is present from the product stockpile and from the landfill.

Scrubber Sludge

Sludge generated at power plants presents a major disposal problem, from both the physical and environmental aspects. The characteristics of sludges from FGD systems vary greatly, depending on the coal burned, boiler, scrubber, reagent used and other factors.

The chemical composition of a sludge is one of the most important considerations in designing a stabilization system, because it can vary greatly, even during standard power plant operation.

One difficulty in developing the stabilization process for a specific power plant is that the scrubber sludge is usually not available prior to the start of actual scrubber operation. As a result, knowledge and experience in stabilizing and handling various types of sludges becomes quite important in establishing the proper process design parameters.

All FGD sludges can be stabilized, but it is important to understand those characteristics of sludge which have the greatest potential effect on stabilization systems.

Sulfite/sulfate proportions primarily affect dewatering. The larger size of the sulfate particles affords easier dewatering. However, given the same ash to sludge ratio, sulfate based sludges require a numerically higher solids content of the final product to be equally handleable than do sulfite based sludges.

Other materials, such as silica, iron, etc. which exist as trace elements in prototype scrubber testing, usually show up as grit in the sludge. Some scrubbers actually produce a cement type material in quantities of 1,000 pounds per day. While that does not affect the chemical stabilization of the sludge, it does affect plant operations and is addressed later in the paper.

The lime or limestone used in scrubbers also have their effects on process design. Poor quality reagent will require that more be added in the scrubber to achieve the required SO₂ removal and the high proportion of non-lime materials increases loads on the dewatering equipment. When in the form of grit, it causes extensive wear on piping and process equipment.

Chemical additives used by the utility to enhance dewatering in the thickener can have significant affect on dewatering. The flocculents which are formed in the sludge to provide increased settling in the thickener are broken down by the shearing action of centrifugal thickener under-

flow pumps and piping systems and cause the vacuum filter medium to become partially blinded. There are several additives available which will provide the necessary thickener settling but minimize filtration problems. These are recommended to the utility where Poz-O-Tec systems are operational.

Solids variations from a thickener are often the result of an upset in scrubber operation or inattention to thickener operation. As thickener solids have a direct relationship to filter cake solids and ultimately final product solids, reduced solids underflow could result in increased disposal costs due to the need for additional handling of the product and the additional quantity of product.

Process Additives

Most stabilization processes require that some sort of additive be used to initiate chemical reactions. Although this activator may already be present in some coals, such as lignite, it must be added separately for most conditions.

For pozzolanic stabilization, the additive most used is lime and it is available as pebble lime requiring crushing, pulverized quicklime, hydrate or lime slurry. The important considerations for the additive lime are (1) CaO content, as that is the necessary activator for the chemical reaction, and (2) lime particle size distribution.

PLANT DESIGN CONSIDERATIONS

Concurrent with the evaluation of process variables to achieve the chemical stabilization, the physical processing systems must also be planned. And, as is true of the power plant and scrubber design, large quantities of materials are involved.

A dry stabilization plant is a materials handling system which processes liquids, sludges, damp solids and dry solids in one facility. Figures 5 and 6 show the stabilization facilities at Columbus & Southern Ohio Electric Conesville Station and Indianapolis Power and Light Petersburg Station.

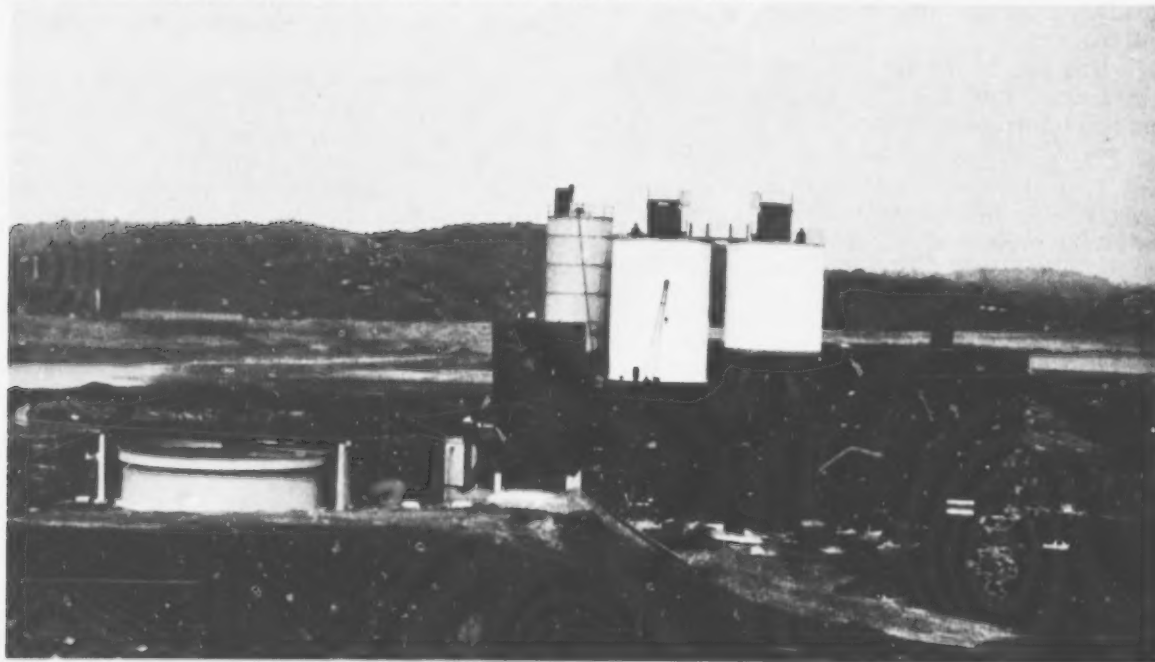
Siting

The location of the stabilization facility will depend on land availability at the power plant, location of the disposal area, and consideration of other factors, both physical and economic.

A location adjacent to the thickeners will minimize pumping and piping costs for underflow and filtrate. Additional savings could be realized if the facility were located next to the power plant ash silos to permit direct feeding of the ash to the facility and eliminate pneumatic conveying between the main silo and surge silo.

Conversely, locating the facility at the disposal area will necessitate extensive transportation systems for liquids and solids. There are also

Figure 5 Columbus & Southern Ohio Electric Conesville Station — SO₂ stabilization facility



limitations for pneumatic ash conveying systems with respect to distance and economics. Pneumatic trucks could be used to transport ash for long distances but that cost would be prohibitively high.

Advantages to locating the facility at the disposal area include reduced transportation costs for the processed material. In some cases, the product could be transported directly by conveyor.

A 600-800 MW plant producing 1,000,000 tons per year of stabilized material will require about 150 acres of disposal area 100 feet high over a 20-year period. Regardless of the plant location and method of product transport to the landfill, the daily delivery of material will normally involve trucks or other hauling equipment.

As can be derived from the above discussion, the most advantageous situation is to have the landfill located near the power plant to benefit from reduced transportation system costs for both the waste materials and the stabilized product. Figure 7 shows the site layout of the IUCS facility at Indianapolis Power & Light Petersburg Station.

Dewatering

Dewatering of the sludge is important in dry stabilization systems because the more liquid removed, the less ash that is necessary to improve handleability and the higher the final product

solids. Some scrubber systems, such as a double alkali system, include thickening and filtration as part of the scrubber operation to reclaim the sodium liquor for reuse. Most FGD systems, however, only thicken the sludge to 25-30% solids.

Dewatering in the stabilization facility is usually accomplished by vacuum filtration, although centrifuges, hydrocyclones and other methods have been considered and used in some limited applications. Figure 8 shows the vacuum filters installed at Columbus & Southern Ohio Electric.

Scrubber sludge can be vacuum filtered at rates of 50-100 lb/sf/hr. and sometimes higher, depending on the chemical composition of the sludge, the filter medium, i.e. cloth, screen, etc. and filter aids. Sulfate sludges usually yield higher filtration rates and solids than sulfites. Excess lime concentrations and sheared flocculents in the sludge may cause cloth blinding.

Filtration will usually dewater a lime based sludge from 30% solids to 42-60% solids and limestone based sludge to 55-60% solids. Oxidized sludges are reported to achieve 80-85% solids, which when mixed with fly ash and additive, would result in a high solids final product. This product may have to have liquid added to achieve optimum placement density. Figure 9 shows effective dewatering ranges for various scrubber sludge compositions.

Figure 6 Indianapolis Power & Light Petersburg Station — SO₂ stabilization facility



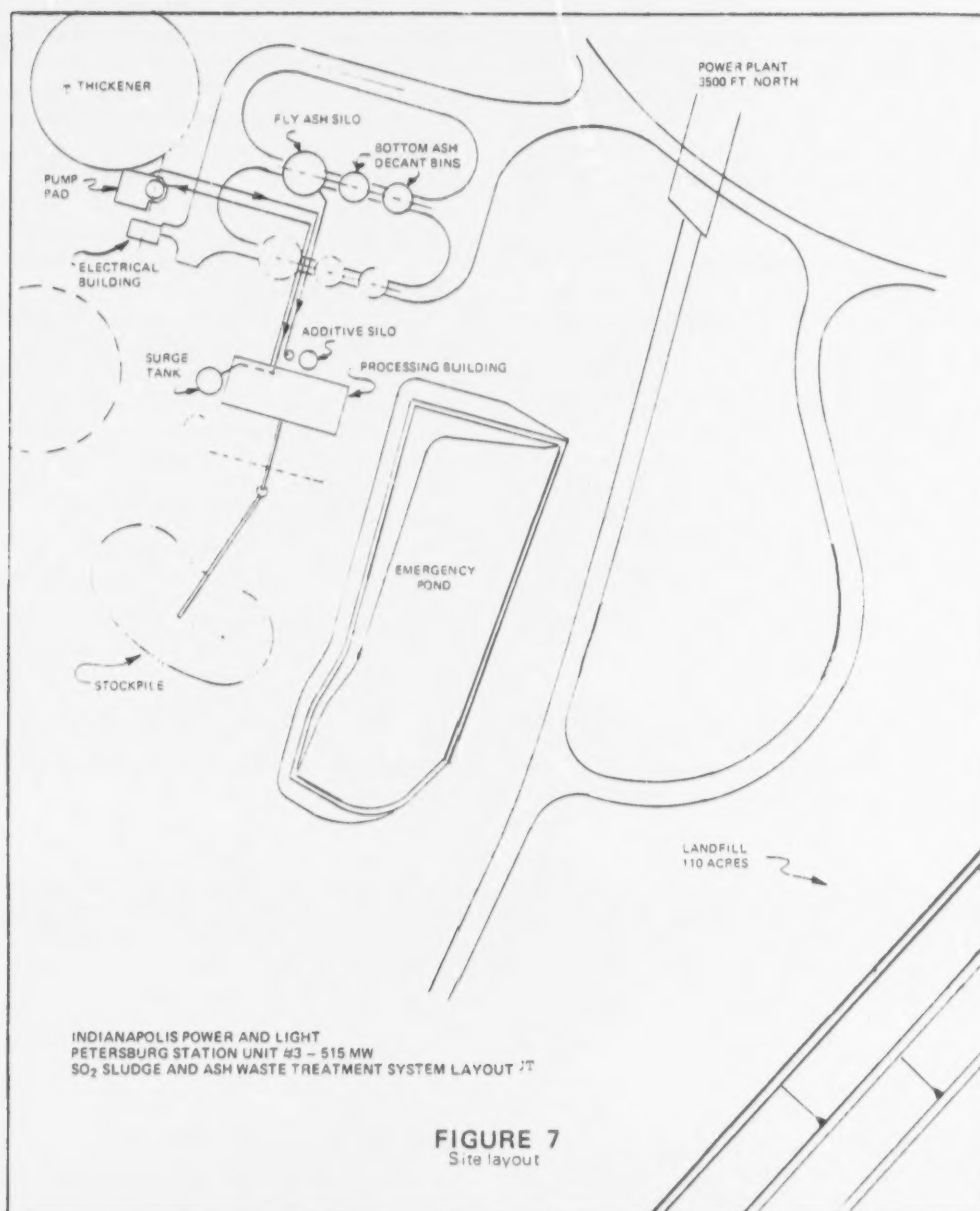


FIGURE 7
Site layout

Centrifuges have been used for dewatering but are more susceptible than filters to variations in sludge composition, solids content and grit.

Hydrocyclones have been offered as a substitute for thickeners and filters but actual field experience is limited. A major drawback to hydrocyclones is the 5% plus solid in the return liquid, which normally exceeds the minimum for the scrubber liquid. Also, hydrocyclones have a critical liquid solids separation phase for which particle sizes below 10 microns are not removable. Most FGD sludges have particle sizes below the 10 mi-

cron size, thereby rendering the hydrocyclone ineffective.

Vacuum filter aides to improve yield and solids can be both mechanical and chemical. Vibrators, compression rolls and flappers can provide significant improvement in filtration. Lime, fly ash and chemical polymers can also provide improvement in filtration when added in the proper proportion. However, too much fly ash in the sludge, a condition which occurs when ash collection equipment is in poor repair, causes blinding of the cloth and actually reduces filtration rates.

Materials Feeding

Feed systems for fly ash and lime involve more than just adding these materials to the sludge. For situations where there is limited ash available, controlled feed is important to conserve ash and ensure that all sludge has ash available for the chemical reaction.

Fly ash is similar to water in that it acts like a liquid, and therefore requires positive control when being fed. Most utilities are experienced in handling ash using collection hoppers, hydrovac systems, and pneumatic conveyors. However, their primary interest is in just moving it, not feeding it at 20-70 tons per hour with a 2-5% accuracy.

IUCS has used several types of ash feeding systems and the two which work best are: 1) a mass flow system; and 2) a gravimetric feeder with a live bin bottom silo. This equipment not only feeds accurately but controls flooding. Those who have witnessed a fly ash flood involving hundreds of tons of fluid ash will appreciate the importance of having a quality fly ash feed system.

As lime constitutes a small percentage of a mix on a dry weight basis and is the activator of the chemical stabilization reactions, accuracy and dispersion of the lime in the product is absolutely necessary. There are a number of feed systems which have the capability to feed the 2-5 tons per hour of lime within the 5% accuracy considered

necessary. Dispersion within the mix, however, depends upon accuracy of feed, location of lime feed into the system, particle size, and quality of mixing.

Mixing

Mixing is the physical combining of the waste and additive material to provide a homogeneous blend to permit the fly ash and additive to contact all sludge particles so complete chemical action can take place.

The mixer must be able to provide the required blending even though the ratio of wet and dry materials may vary over any given period and the mixer designed for 200 TPH may only be operating at 100 TPH due to reduced station load.

The specific combination of waste materials to be mixed at a facility must be evaluated for material ratios, solids content, particle size, retention time, type additive, etc. to ascertain the proper mixer design.

The most difficult material to handle in a mixer is filter cake. It clings to the mixer housing and paddles; it may not blend with the other materials if not mixed for the proper duration. If mixed too much, it will become thixotropic and even with the ash mixed in, a semi-fluid material will exit the mixer.

Figure B Vacuum filters — Columbus & Southern Ohio Electric — Conesville plant

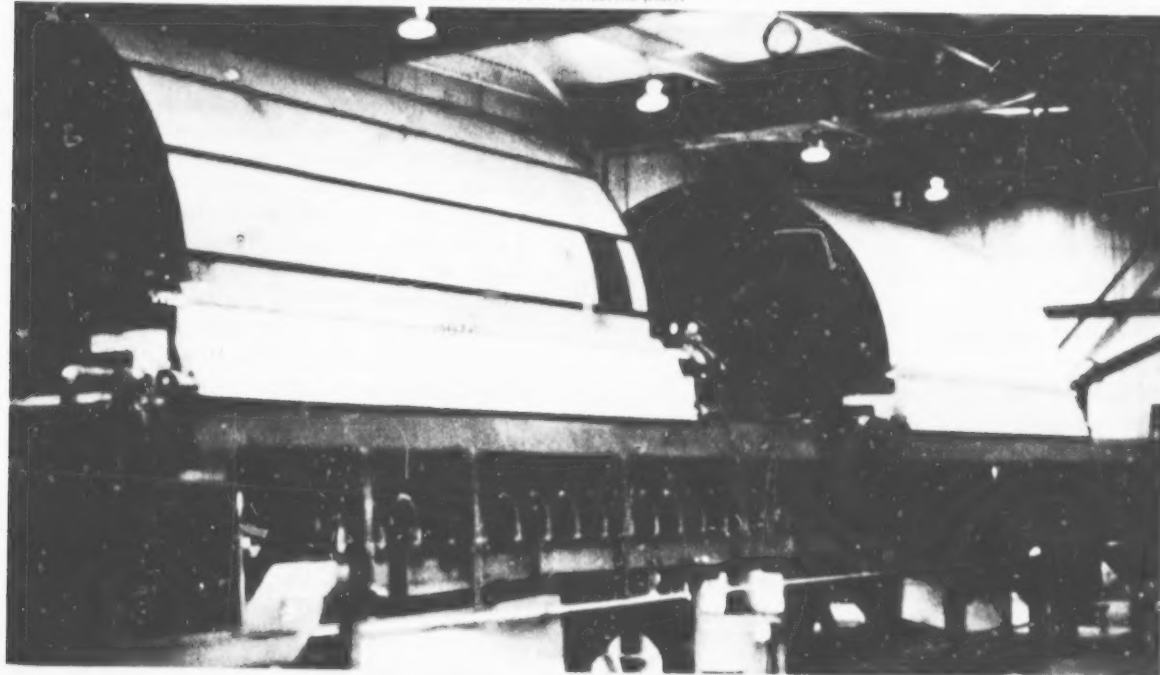


FIGURE 9
SLUDGE DEWATERING CHARACTERISTICS

SCRUBBER	SLUDGE COMPOSITION	EFFECTIVE DEWATERING RANGE
Magnesium Modified Lime	$\text{CaSO}_3/\text{CaSO}_4/\text{MgSO}_x$	42% - 50%
Lime	$\text{CaSO}_3/\text{CaSO}_4$	50% - 60%
Limestone	$\text{CaSO}_3/\text{CaSO}_4/\text{CaCO}_3$	55% - 60%
Dual Alkali	$\text{CaSO}_3/\text{CaSO}_4/\text{Na}_2\text{SO}_4$	50% - 60%
Magnesium Modified Lime with partial fly ash removal	$\text{CaSO}_3/\text{CaSO}_4/\text{MgSO}_x$ + fly ash	48% - 52%

With the IUCS system approach, the reactive chemistry is such that variation in product consistency can be tolerated and a satisfactory final product still achieved.

Much of the equipment in the stabilization facility is process related, such as dewatering, feeding and blending. Other equipment is supportive, but of equal importance in designing a system. Two of these, dust collectors and conveyors, require special attention and will be discussed briefly.

Dust Collectors

IUCS has a number of dust collector systems on its plants, and experience has shown that these systems pass some dust. Such things as particle size, grain loadings, humidity, dew point, air to cloth ratio, and quality and quantity of purge air have a significant effect on dust collector performance.

Dust collectors for mixers where wet and dry materials are combined also warrant special consideration. Depending on the additive used, the mixing operation can produce a steam which hinders dry dust collection. In some cases, a wet scrubber is necessary.

Conveying

Belt conveyors are used in stabilization plants to transport filter cake to the mixer, damp fly ash or bottom ash to the mixer, final product to a stacker and for the stacker itself.

Dry or slightly damp materials are not much of a conveying problem. Wet materials, especially those having thixotropic properties, are another matter. That type material clings to belts, scrapers, transfer points, and has to be physically removed if the conveyor system is not properly designed.

It is important that all components of a belt conveyor system be designed together. To install a

"super" belt cleaner without providing adequate support against which the cleaner can act, will result in a non-cleaned belt. To design a transfer point without an adequate hopper will result in a plugged transfer point or a pile of material on the ground. This may seem basic, but the knowledge comes from actual experience, not from the manufacturer of the equipment. Figure 10 shows a typical belt scraping system.

Landfill considerations must also be taken into account when designing a stabilization facility. Radial stacker capacity depends on hourly production rates, consistency of the product, whether interim conditioning is required before placement, the disposal schedule and the equipment planned for the transport of the product to the landfill. Figure 11 shows the radial stacker installed at Indianapolis Power and Light Petersburg Station.

PROCESS OPERATION AND MAINTENANCE — SOME EXPERIENCES

IUCS has four stabilization facilities operational which involve a representative selection of scrubber sludges, ash, additives, landfills and other operational aspects. Scale-up of both process and design considerations has been a learning process and sometimes a difficult and expensive one. One thing is certain: Scale-up from a 100 pound bench scale mix to a 200 ton per hour, 24 hours per day, 7 days per week operation is a significant step and IUCS has gained considerable experience by operating four plants for periods up to four years. Three of these plants are owned and operated by IUCS and this has become an excellent vehicle by which to increase our knowledge of plant design requirements.

Much of what IUCS has learned in operating plants is used to upgrade existing plants and those presently under design or construction. These improvements are also available to utilities who are operating IUCS stabilization facilities.

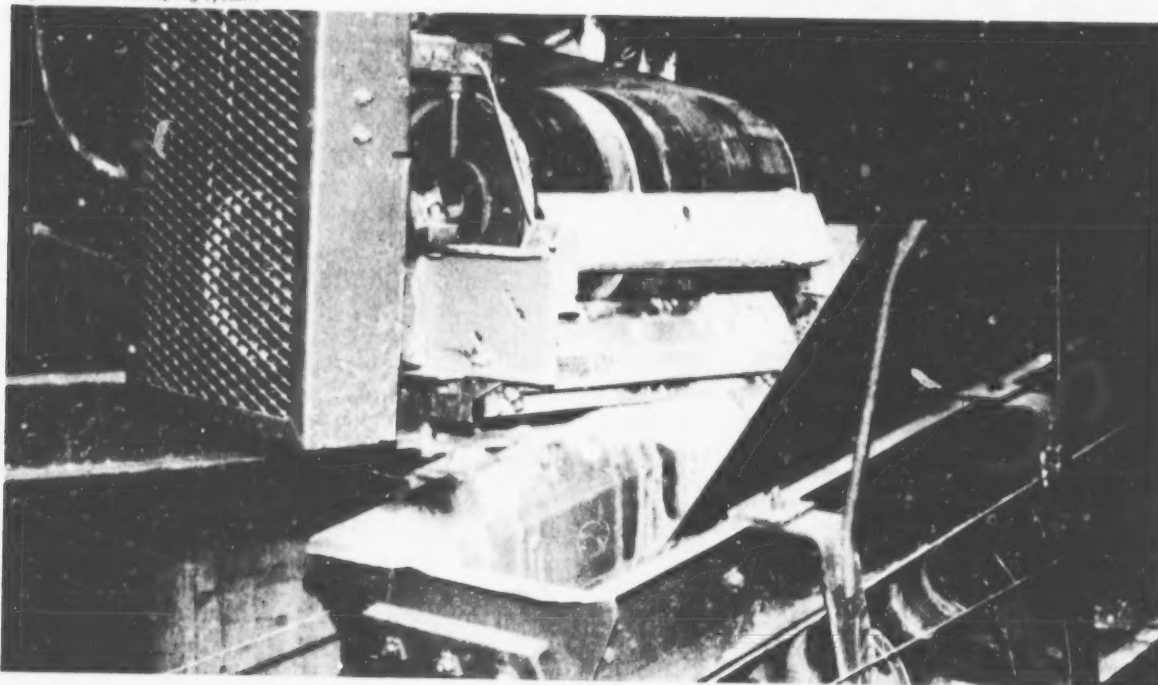
Start-up of stabilization plants, although not as complicated as a scrubber start-up, is still important since the stabilization facility must produce a product that will satisfy environmental regulations and allow continuous scrubber operations.

Start-up consists of (1) equipment checkout, (2) determination of equipment capability now that the actual sludge and ash are produced and (3) confirmation of the expected chemical reaction. Operating personnel must be trained on the equipment and quality control procedures.

Quality control is necessary on a continuous basis as a tool to ascertain what is being received, what should be expected when wastes are pro-

cessed, what is actually produced and placed and what process modifications are necessary when variations occur. Therefore, the initiation of the quality control program concurrent with the equipment startup is a necessity.

Figure 10 Belt scraping system



Actual operational experiences can best be described by reviewing some instances following the flow of materials through a stabilization facility starting with the scrubber sludge. Earlier, a discussion on grit in the sludge was presented.

Figure 11 Radial stacker — Indianapolis Power & Light Petersburg Station



This grit causes pipes to wear out, especially unlined pipes. Grit wears heavily on rubber lining of pumps and valves causing replacement in less than a year when 5 years life was expected. It settles in pipe joints, tees, and elbows plugging them as the "hybrid cement" hardens. It builds up in surge tanks causing agitators, which were designed for 30% suspended solids, to seize up. This problem is compounded in thickeners. Several utilities, as well as one of our own plants, have experienced this problem. Removal of four feet of settled and possibly reacted sludge/cement in a thickener is time consuming and expensive.

For the most part, Utilities operate thickeners with a primary concern for overflow clarity, since this liquid is used for scrubber makeup. Thickener underflow usually receives less attention. Underflow is usually specified contractually at a certain percent solids, however in actual operations, solids concentration vary widely. For example, at one installation 30% solids at 300 GPM was specified, although the stabilization plant received 20% solids at 450 GPM. The lower solids produce lower filtration yield and, at the higher than expected flow rate, the system surge tank level increases to a point where a back-up filter needs to be operated. When that occurs, the filtrate load exceeds the sump and sump pump capacity and the cake may exceed the conveyor and mixer capacity.

Excess lime in the scrubber may be necessary to achieve compliance with emission requirements or to minimize scaling. However, that excess lime passes through the thickeners and affects the filtration operation. The lime is usually fine particle material, blinds the cloth, and significantly reduces the filtration rate. Acid washing of the cloth has been tried, and although somewhat effective, it has met with resistance from the operating personnel who actually have to handle the acid. Older boilers may use oil during combustion and that also blinds filter cloth.

One answer to these problems is frequent filter cloth changing. Given experienced personnel, a complete cloth change on a 12' diameter x 20' length vacuum filter can be accomplished by 3 men in eight hours. Labor and material costs per change is \$800-\$1,000 depending on labor rates and cloth type. Changing cloths weekly, as has been done in one of the stabilization plants, cost \$50,000 per year. The life cycle for filter cloth was expected to be three months per the filter manufacturers experience.

Temperature is a factor in vacuum filtration. Sludge of higher temperatures will yield significantly higher filtration rates than at lower temperatures. When a utility installs several thousand

feet of uninsulated sludge pipe line four feet off the ground in areas where the temperature and wind chill factor are frequently below zero degrees F, sludge temperature will drop considerably from the utility's thickener to the processing facility. Hence, consideration should be given to reheating sludge prior to filtration as an alternative to additional filters to achieve the necessary production.

Filtration can be aided by the addition of lime, at certain particle sizes, or fly ash in the sludge. Improvements of 50% in filtration rates have been obtained in pilot tests and full scale testing is being conducted.

Previously, the importance of fly ash feed was discussed. At C&SOE's Conesville Station, IUCS installed vibrating pan feeders and live bin bottoms on a 500 ton capacity silo. The vibration caused cracking of the vibrating motor bracket attached to the bin bottom without providing controlled ash feed. It did cause considerable ash flooding through a 150 foot screw conveyor into and out of the mixer, into the processing building and onto the discharge conveyors — several hundred tons, at times. Experience has shown the importance of good ash feed equipment to prevent floods. Mass flow feed systems have now been installed and consistent accurate feed has been obtained.

Exterior belt conveyors handling damp materials have also presented challenges. Visualize a 24" belt conveyor on 30° troughing idlers having several transfer point hoppers with 45° sides, simple rubber scrapers and discharging onto a stacker. Approximately 80% of the material will drop off the belt at the head pulley by gravity. The balance is carried to the belt scraper which removes another 18%, leaving 2% on the belt. For a 150 ton per hour plant that 2% is equivalent to 74 tons of material per day which could plug pulleys, idlers, rollers, and then have to be picked up by shovel and wheelbarrow. IUCS is utilizing several types of belt cleaners, sometimes in tandem, to keep the belts clean. A good belt cleaning system has to be 100% effective to be good.

The 98% of the material removed from the belt is discharged into the transfer hopper and, because the material is wet, it adheres to the 45° steel sides usually resulting in bridging and spillage. Transfer hoppers with 80°-90° sides are required to prevent this problem. At the radial stacker hopper, spillage has caused additional problems. Some of the material drops at the pivot point and unless cleaned regularly, it will set up or freeze. If the latter occurs, and the stacker is rotated, it rides up the hard material and separates from the pivot point. Result: an inoperable radial stacker system and, if no backup is available, an inoperative plant.

To prevent such problems, plant design, equipment redundancy, operating philosophy, and maintenance programs must be planned, coordinated, and evaluated.

For plants which IUCS operates, maintenance capability is provided on all operating shifts to minimize backup equipment for certain process systems. For turnkey plants, experience has shown that utilities prefer to have a high degree of redundancy and backup equipment, and limit maintenance to day shift operations. Since the stabilization systems are often located remote to the immediate power plant area and since problems may occur simultaneously with turbine, boiler, or scrubber equipment, stabilization systems have characteristically received a low priority of maintenance.

Maintenance requirements vs. backup equipment is a compromise and must be addressed by plant operations as well as engineering. This can be accomplished by good planning up front with due consideration to capital expenditure requirements and power plant operation philosophy.

FINAL PRODUCT HANDLING

The achievement of a structurally stable and environmentally compatible landfill requires a detailed material handling and placing program, landfill preparation, and quality control procedures. In many respects, the disposal and placement procedures are as important to the overall stabilization system as the processing facility itself. The overall treatment system involves two areas of consideration: the processing plant and the landfill.

The operating and design parameters of the processing facility have been discussed at length. However, it is important to understand that the methods and techniques used to handle the processed material are also of key importance to the overall system design.

Once the processed sludge leaves the facility it is normally placed in a surge pile. Time in the surge pile can be for a couple of hours to 4-5 days depending on the physical and chemical characteristics of the sludge, availability of fly ash, final product solids, and disposal schedule.

Normally, a dryer consistency final product will require shorter time in the surge pile prior to handling. For example, final product solids in the 70-75% range can be moved almost immediately, whereas product solids range of 50-58% will require initial conditioning of several days before movement. Figure 12 gives a range of final product solids and required conditioning times prior to movement.

FIGURE 12

Solids Range (%)	Stockpile Conditioning (Days)	Stockpile Capacity (Days)
50-58	4+	8
58-63	3-4	7-8
63-70	1-3	5-7
70-80	0-1	4-5

If the material is moved too soon after production, it tends to exhibit thixotropic tendencies requiring greater handling care. The material is difficult to place in the landfill as it becomes "oozy" and tends to stick to and inhibit movement of dozers and compaction equipment. The material can be quite slick requiring special material handling equipment to affect efficient placement.

Temperature can also affect the material cure, handling and placement operations. During winter months, with temperatures below 40°F, the chemical reactions in the material are slowed — similar to cement chemistry. As a result, greater curing times may be required for the material in the surge pile prior to movement and placement in the landfill. Adequate storage capacity in the surge pile area must be accomplished during system design to plan for this requirement.

The processed material can be moved to its final disposal area by (1) front end loading directly onto trucks; (2) by direct conveyor to the landfill; (3) by unit train to landfill area, etc. In all instances grading, placing and compaction is required for final placement. The optimum mode of transportation is based on site-specific parameters, i.e., distance to landfill, landfill access, product characteristics, and Utility preferences. The attached Figure 13 shows typical processed material being placed in a landfill.

Material is usually placed in 12 to 24" lifts to ensure adequate placement density and curing. The disposal operation should be maintained such that a minimum surface area of fresh material is exposed to the elements. The working face should have a slight grade such that any rainfall will tend to runoff rather than collect in pockets. Should rain water pockets occur, especially on fresh material, the material stabilization will be adversely affected or stopped completely creating soft spots in the landfill. It is very difficult for material handling equipment to move on or through this material.

In the landfill plan, the material can be built-up to elevations up to 100' and greater. The landfill is developed in approximately 25' lifts and benched at the outer surface to provide haul roads and prevent erosion. Side slopes can be 2:1 horizontal to vertical with 50' benches. The finished surfaces should have at least an 18" layer of top soil and be revegetated to retard erosion.

At several of IUCS' installations, long range plans call for material to be built into mountains in excess of 100' in height thus minimizing land area requirements. Normally, these landfill areas are on or adjacent to utility property so that a minimum of material hauling is required.

The biggest potential environmental impact could be caused by water runoff from the landfill. For this reason, exposed surface area of freshly placed material should be kept to a minimum. IUCS has used sedimentation ponds to collect the runoff discharge from the landfill area. Suspended solids tend to settle out with the supernatant suitable for discharge to the terrain or evaporation.

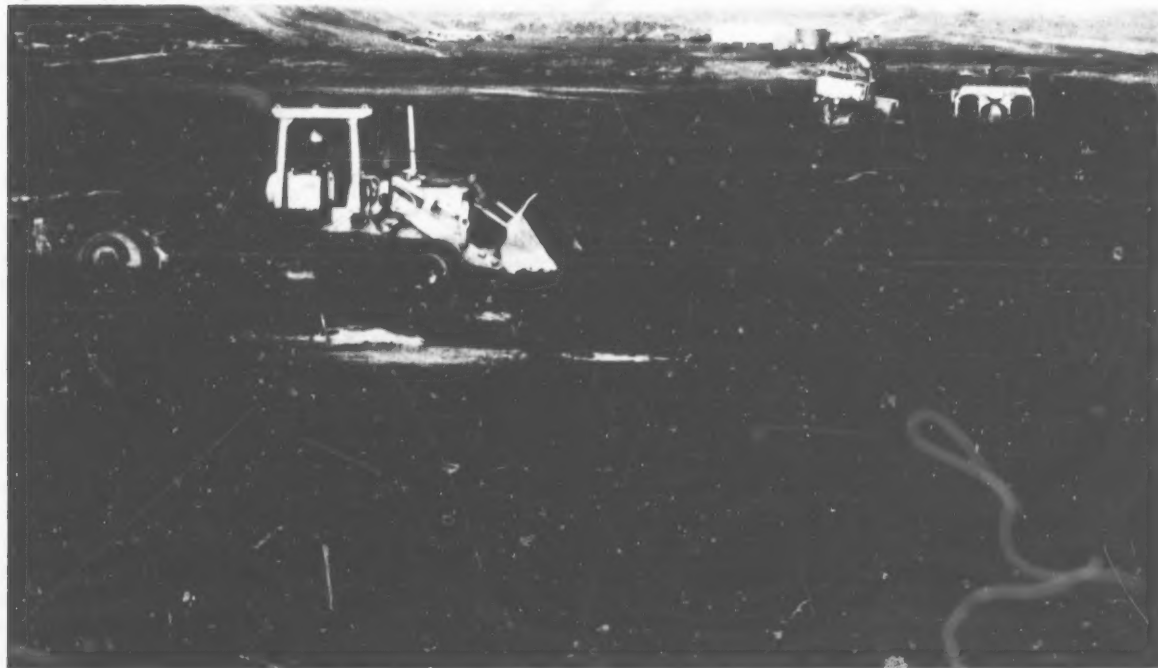
A good landfill plan and operation will require the use of monitoring wells to sample ground water. These wells should be installed well in advance of the beginning of operations to obtain appropriate background data.

REGULATORY OVERVIEW

Regulatory approval has not presented a problem to date. Since most disposal sites are on utility property, the utilities must apply for the necessary permits. While IUCS stands ready to support all its clients in the permitting procedure, most utilities prefer to handle the matter themselves. However, IUCS takes positive steps to smooth the way for permit approval. IUCS attempts to obtain conceptual approval and to provide an understanding of our process in all states in which we anticipate business before bid specifications are issued. The regulatory agency is therefore familiar with the process before the permit application is submitted. To date, all permit applications have been approved including two submitted by IUCS for landfills that we operate. The system has been accepted in the following states: Indiana, Illinois, Kentucky, Ohio, Pennsylvania, Texas, Florida, West Virginia and New Mexico.

Since no one knows, at this time, what the RCRA regulations will be for fly ash or scrubber sludge — will they be classified as hazardous or non-hazardous — IUCS has been working with the Federal Office of Solid Waste to demonstrate the benefits of a structural landfill constructed of materials with good engineering properties. The

Figure 13 Typical stabilized material placement



outcome of our efforts will not be known until the requirements for special wastes have been defined later this year. IUCS believes that it has been able to demonstrate that a properly designed and managed structural landfill of stabilized sludge/ash provides equal or better environmental protection than procedures permitted in the proposed regulations. If the EPA agrees with our calculations, there would be no increase in cost to the utilities for disposal even though the sludge or ash is classified hazardous.

OPERATIONAL DATA

The operation of the four dry stabilization systems discussed in this paper has produced a total of 1,627,000 tons of stabilized material. In 1983, when the additional eleven stabilization facilities are operational, 22,491,060 tons per year of stabilization capacity will be available to assist in minimizing environmental pollution.

Design, production and landfill data for the four IU Conversion Systems commercially operating stabilization systems is shown on Figure 14.

FIGURE 14
SPECIFICATION DATA – OPERATIONAL STABILIZATION SYSTEMS

Utility	Plant	MW	Scrubber	Design Capacity Tons/Hr.	Annual Production 100% L. F. Tons/Hr.	Operational Date	Total Production To Date Tons/Hr.	Disposal Services
Duquesne Light Co.	Phillips	400	Magnesium modified lime	70	613,000	11/77	489,000	Total disposal services including permitted site
Duquesne Light Co.	Elrama	500	Magnesium modified lime	100	876,000	11/75	833,000	Total disposal services including permitted site
Indianapolis Power and Light Co.	Petersburg #3	515	Limestone	130	1,138,000	9/78	37,000	Design and technical consultation. Utility site adjacent to plant
Columbus & Southern Ohio Electric Co.	Conesville #5 & #6	860	Magnesium modified lime	170	1,490,000	1/77	268,000	Design & management. Utility site adjacent to plant

SUMMARY

SO₂ scrubber sludge and ash disposal has become an important consideration in the design of power plants. In the stabilization process by IUCS, fly ash is blended with dewatered sludge and lime additive to form a product which is suitable for landfill disposal. Chemical and physical characteristics of the waste products must be analyzed and their affects known to optimize the process design.

The stabilization system design must include an analysis of the optimum siting location. The availability of land at the power plant, location of disposal area, and material transportation analysis must be accomplished.

Dewatering of sludge is important in a dry stabilization system as the material handleability and chemical reactions will be affected. Sludge dewatering is dependent on the chemical composition of the sludge, i.e. sulfate to sulfite ratio, excess lime concentrations, magnesium concentration, etc. and the physical characteristics, i.e. particle size and sheared flocculants.

Material feeding and blending must be controlled to insure optimum final product characteristics. This is especially true when limited fly ash is available to the process. Lime feed accuracy is important so that additive costs are minimized. Blending of the material requires special techniques such that the system can accept a wide range of wet and dry material. Over mixing will create thioxotropic final material and under mixing creates poor dispersion of individual materials resulting in degradation of structural and environ-

mental final product characteristics. Dust collection equipment should be used which takes into consideration blending and conveying of wet and dry materials. Conveyor and radial stacking requirements must include an analysis of material quantity, quality, handleability, tendency to fluidize, and rate of chemical stabilization reaction time.

IUCS has had considerable process operation and maintenance "hands-on" experience based on four stabilization systems in operation. All of the systems have been operating since startup without a major outage, however, several problems have been encountered from time to time. These problems have included sludge pipe pluggage and wearout because of grit in the sludge, controlled fly ash feeding problems, sludge dewatering challenges, vacuum filter cloth blinding, conveyor belt cleaning difficulties, material handling challenges etc. With the state-of-the-art development in this field, and the many problems that can be encountered, IUCS has recognized the need for an intensive quality control program to ensure smooth and continuous operations.

The regulatory outlook for power plant wastes is still in question at this time. IUCS believes, however, that it has been able to demonstrate that a properly designed stabilization system with a properly managed landfill will yield a final product and long term solution to power plant waste disposal which is environmentally sound and economically feasible.

POWER PLANT FLUE GAS DESULFURIZATION USING
ALKALINE FLY ASH FROM WESTERN COALS

by

Harvey M. Ness,¹ Philip Richmond,²
Glenn Eurick,³ and Rick Kruger⁴

ABSTRACT

A characteristic of Western coals is that they contain high levels of alkali such as calcium, magnesium, and sodium. These alkali species can be leached from power plant fly ash for use in flue gas desulfurization (FGD) wet scrubbers in lieu of lime or limestone. At present, there are nearly 2,600 MW's of generating capacity in the Western United States that utilize either fly ash or fly ash supplemented with lime or limestone. An additional 3,500 MW's of Western generating capacity are being planned or are in various stages of construction which will use the alkaline fly ash. This report describes the Western alkali ash FGD systems.

¹ U.S. DOE, Grand Forks Energy Technology Center, Grand Forks, N. Dak.

² Square Butte Electric Cooperative, Center, N. Dak.

³ Minnesota Power and Light Company, Duluth, Minn.

⁴ Northern States Power Company, Becker, Minn.

POWER PLANT FLUE GAS DESULFURIZATION USING ALKALINE FLY ASH FROM WESTERN COALS

INTRODUCTION

Two important characteristics of most Western coals is that they contain significantly less sulfur than most Interior and Eastern region coals, and that the coal ash contains high levels of alkali such as calcium, magnesium, and sodium. The significance of the generally low sulfur content, which averages about 0.7 pct, is that a sulfur dioxide removal efficiency of only 30 to 40 pct is required to meet the existing Federal emission standard of 0.52 g/MJ (1.2 lb SO₂/MM Btu), although higher removals are required to meet some state and local standards. The lower removal requirements, however, have allowed novel developments in stack gas cleanup technology.

The alkaline constituents in Western coals tend to be the highest in the low-rank lignites, and progressively less prevalent in the subbituminous and bituminous coals. The alkaline oxides in the ash can vary widely, ranging from under 10 pct to over 50 pct, with significant variations from mine to mine in a specific coal region, and even between locations within a single mine. Characteristic differences in coal ash compositions are illustrated in Table 1.

Table 1. ASH COMPOSITION RANGES FOR WESTERN
BITUMINOUS AND LIGNITE COALS¹

	Bituminous, pct	Lignite, pct	
		North Dakota	Texas
Silica, SiO ₂	20 to 60	11 to 28	38 to 66
Aluminum oxide, Al ₂ O ₃	10 to 35	8 to 14	14 to 23
Ferric oxide, Fe ₂ O ₃	5 to 35	2 to 16	1.8 to 11.8
Calcium oxide, CaO.....	1 to 20	18 to 31	3.8 to 20.5
Magnesium oxide, MgO.....	0.3 to 4	2 to 9	0.9 to 6.1
Sodium oxide, Na ₂ O.....	0.2 to 3	1.4 to 6.5	0.2 to 4.5
Potassium oxide, K ₂ O.....	0.2 to 4	0.2 to 0.5	0.1 to 1.8
Titanium oxide, TiO ₂	0.5 to 2.5	0.2 to 0.6	0.5 to 1.6
Phosphorous pentoxide, P ₂ O ₅ ..	0.0 to 3	0.0 to 0.6	0.0 to 0.2
Sulfur trioxide, SO ₃	0.1 to 12	12 to 27	2.6 to 19

The potential usefulness of the alkali in Western coal ash for flue gas desulfurization can be illustrated by comparing the mole ratio of coal ash calcium oxide to coal sulfur content. For a coal containing 7.5 pct ash, with 25 pct calcium oxide in the ash, the calcium oxide is chemically equivalent to slightly more than 150 pct of a 0.7 pct coal sulfur content. The alkali in some lignites can have an alkali-to-sulfur mole ratio of several hundred percent. The amount of fly ash alkali that would be available for use in a wet scrubber would depend on the coal ash content, method of firing (pc versus cyclone), and scrubber operating conditions; however, there is often ample fly ash alkali available to react with sulfur dioxide in a wet scrubber.

Laboratory studies have shown that the fly ash alkali can be solubilized into an aqueous media, and that solubilization is primarily a function of pH and, to a lesser extent, mix time. Figure 1 illustrates the calcium oxide availability as a function of pH for three North Dakota lignite fly ashes. The data were generated using batch leach procedures² and, in general, indicate an increase in the amount of available calcium oxide as the pH of the solution decreases. Similar trends are evident for most Western coal fly ashes. There are, however, differences in leach characteristics for fly ashes from different mines and power plants. The calcium oxide content of the three fly ashes shown in Figure 1 are nearly identical; however, the amount of available calcium oxide at pH 7 varies from about 10 pct to about 40 pct. As the solution pH decreases, however, the differences in solubility become less significant and tend to approach similar availability values. The variations in alkali solubilities can be attributed to differences in chemical compositions of the original coals, and also to differences in the boilers from which the fly ash was derived. Other major fly ash species, such as magnesium and sodium, can also be leached. Magnesium exhibits a pH profile similar to calcium, while the amount of available sodium is not significantly affected by pH. The alkali availability data indicate that significant amounts of calcium alkali can be leached from fly ash.

FULL-SCALE FGD UNITS USING ASH ALKALI

As noted previously, the relatively low sulfur content combined with the high levels of soluble alkali in the Western coal fly ash have resulted in some innovative engineering for flue gas desulfurization (FGD) wet scrubber systems on Western power plants. Three extensive pilot plant programs investigating fly ash alkali wet scrubbing have been conducted, and have resulted in the construction of full-scale scrubber systems utilizing fly ash alkali. A fourth pilot plant test program is currently being conducted. Additionally, the Department of Energy's Grand Forks Energy Technology Center (GFETC) has been investigating the use of alkaline fly ash in a pilot plant wet scrubber since 1970.

At present there are nearly 2,600 MW's of generating capacity that utilize fly ash, or fly ash supplemented with lime or limestone. An additional 3,500 MW's of generating capacity that will use all or part of the alkaline fly ash are currently being planned or are in various stages of construction. The FGD units that are presently in-service vary in design from venturi-marble bed, to venturi-spray tower, to only a spray tower

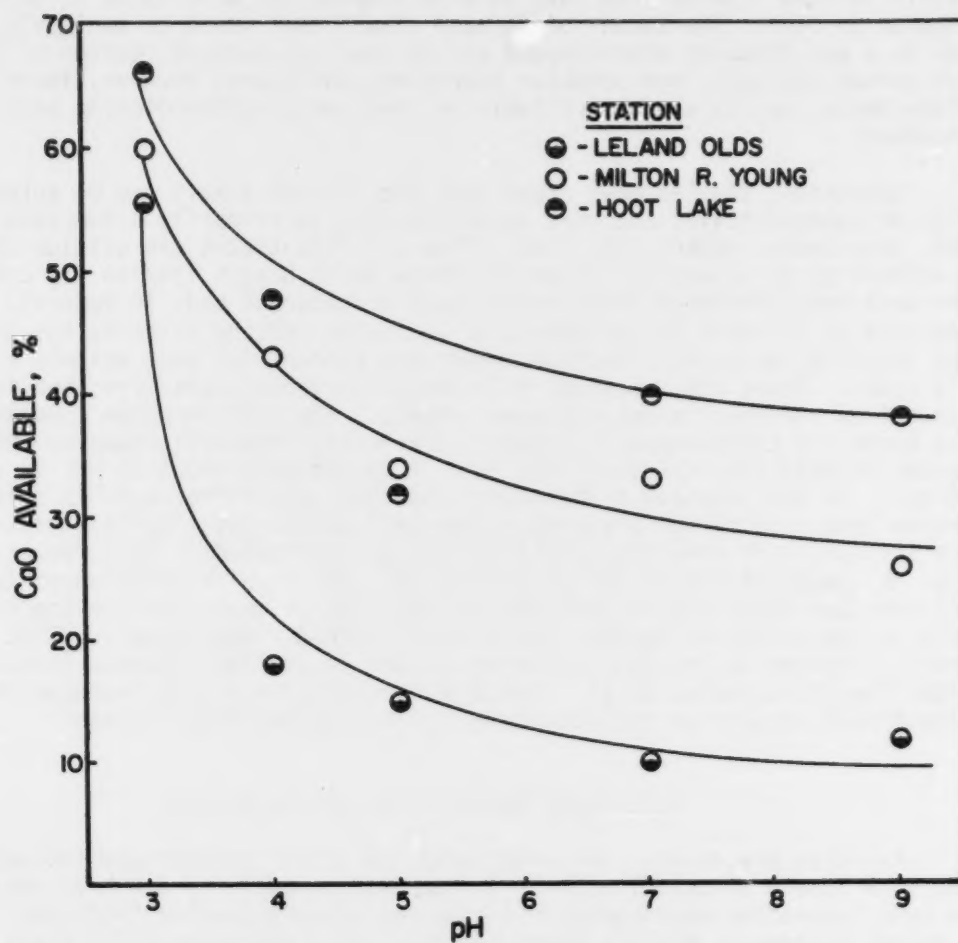


Figure 1. Effect of pH on calcium availability for three Western coal fly ashes.

preceded by an electrostatic precipitator. A summary of the current utility pilot plant programs, along with planned and operating fly ash alkali FGD wet scrubbers, are presented in Table 2. A brief description of each installation is presented below.

NORTHERN STATES POWER COMPANY (NSP)

NSP is currently operating two 700-MW boilers which were put into service in May 1976 (Unit 1), and May 1977 (Unit 2) at the Sherburne County Generating Station. A third unit is planned to be in operation in 1984. The generating units are required to meet a permit requirement of 0.037 g/MJ (0.087 lb/MM Btu) for particulate (99 pct removal), and 0.41 g/MJ (0.96 lb/MM Btu) for sulfur dioxide (about 50 pct removal). The primary fuel is a subbituminous coal from the Colstrip area of Montana and has the following typical analysis: 25 pct moisture, 9 pct ash, 0.8 pct sulfur, and a HHV of 19,733 J/g (8,500 Btu/lb). A typical fly ash would contain 17 pct calcium oxide.

A pilot plant test program investigating the utilization of fly ash alkali was conducted by NSP, Combustion Engineering, and Black and Veatch Consulting Engineers. A conclusion of the test program was that fly ash alkali could provide 70 to 80 pct of the alkali requirements, with the remainder added as limestone.³ Based on pilot plant data, two-stage scrubber systems were installed on Units 1 and 2.

A schematic of the flue gas cleaning system is shown in Figure 2. The system was designed to remove particulate in a venturi, and sulfur dioxide in a marble bed. The L/G at the venturi is about 17, and the L/G at the marble bed is about 10. Each system consists of 12 scrubber modules; 11 modules are required for full-load boiler operation. Each module consists of a venturi section, a marble bed, a mist eliminator section, and a reheater section.

The section containing the marble bed is constructed of carbon steel coated with flakeglass lining for corrosion protection. The slurry reaction tank is not coated with flake fiberglass lining, and provides a liquor holdup time of about 12 minutes. A single pump with Ni-Hard internals provides slurry to the venturi and the marble bed. As solids accumulate in the module, part of the slurry is bled to a thickener, and the underflow is pumped to an ash pond. Cooling tower blowdown water is added to the thickener overflow and ash pond liquor, and is used for: 1) mist eliminator wash water; 2) flush water for slurry lines; and 3) makeup water to the scrubber modules. The scrubber systems are operated in a closed-loop mode; no liquor is discharged to surface waters.

The flue gas particulate is removed in the venturi section using slurry from the reaction tank. The fly ash trapped in the water droplets drains to the slurry mix tank when the calcium oxide leaches and becomes available for reaction with sulfur dioxide. The calcium available in the fly ash (5.7-6.9 g/m³ [2.5-3 gr/scf], 17 pct CaO) represents approximately 60 to 70 pct of the total alkali put into the scrubber system. Supplemental calcium is added as limestone at a rate of about 2.7 metric tons/hr (3 tons/hr). The limestone is ground in wet ball mills and delivered to the

Table 2. FGD WET SCRUBBERS UTILIZING ALKALINE WESTERN FLY ASH

Utility/Station	Capacity	Status	FGD Design	Alkali Source
Northern States Power Co. Sherburne County	Unit 1 - 700 MW	In-service	Venturi-marble bed Venturi-spray tower	Subbituminous fly ash - limestone
	Unit 2 - 700 MW	In-service		
	Unit 3 - 700 MW	Planned -- 1984		
Minnesota Power & Light Co. Clay Boswell	Unit 3 - 1 MW	Pilot plant	Venturi-spray tower	Subbituminous fly ash - lime
	Unit 4 - 500 MW	Under construction		
	Unit 1 - 500 MW	Planned -- mid-1980's		
Montana Power Co. Colstrip	Unit 1 - 360 MW	In-service	Venturi-spray tower	Subbituminous fly ash - lime
	Unit 2 - 360 MW	In-service		
	Unit 3 - 700 MW	Planned		
	Unit 4 - 700 MW	Planned		
Square Butte Electric Coop. Milton R. Young	Unit 2 - 440 MW	In-service	Spray tower	Lignite fly ash - lime
United Power Association- Cooperative Power Association Coal Creek	Unit 1 - 550 MW	Under construction	Spray tower	Lignite fly ash - lime
	Unit 2 - 550 MW	Under construction		

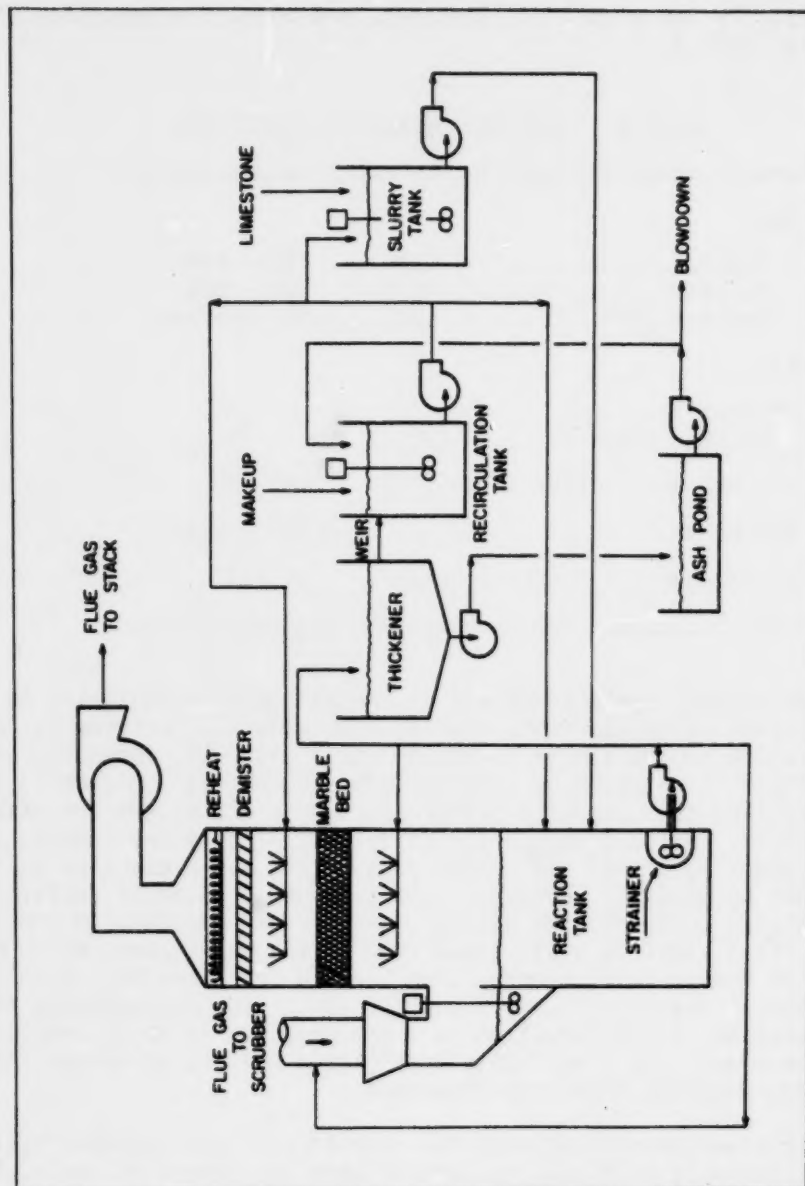


Figure 2. Flow schematic for Sherburne County FGD system, Northern States Power Co.

slurry reaction tanks as a 4 pct slurry. The limestone represents about 30 to 40 pct of the total calcium added to the system and is used to maintain the slurry pH at a range of 5.0 to 5.5. The average sulfur dioxide removal efficiency is about 60 pct (inlet concentration range of 500-650 ppm). The fly ash calcium reacting with the absorbed sulfur dioxide represents about 50 pct of the total amount available. A summary of the sulfur dioxide removal efficiencies, fly ash alkali utilizations, and operating conditions is presented in Table 3.

Table 3. NSP FGD OPERATING CONDITIONS

SO ₂ :		
Inlet.....	ppm..	500 - 650
Outlet.....	ppm..	200 - 300
Removal.....	pct..	60 (average)
L/G:		
Venturi.....		17
Absorber tower.....		10
Fly ash utilization.....	pct..	40 - 60
Slurry pH.....		5.0 - 5.5
Limestone.....	metric ton/hr..	2.7

Calcium sulfate scale formation is controlled by circulating 10 pct suspended solids in the scrubber slurry. The amount of calcium sulfate crystals present in the slurry is about 2 to 3 pct. The formation of calcium sulfate is enhanced by forcing about 350 pct stoichiometric air into the reaction tank, and the sulfur species in the system are essentially 100 pct sulfate. Without forced-air oxidation, pilot plant results indicated that about 40 pct of the sulfur in solution was present as sulfite, and the remainder as sulfate. However, sulfite would oxidize to sulfate, and since the slurry contained few calcium sulfate seed crystals to provide nucleation sites, sulfate scale would form on the marble bed, which would result in the module being removed from service for cleaning. With forced air oxidation in the slurry reaction tank, which provides adequate time for desupersaturation, scale formation on the marble bed is not a problem. Bed pluggage, however, resulting from plugged spray nozzles or uneven flow distribution, requires regular maintenance.

Past problems associated with the chemistry of the scrubber system which have been most difficult to control were the levels of recirculated solids and slurry pH. Other problems have been excessive wear of spray pumps, failure of rubber-lined spray piping, pluggage of strainers on the spray pumps, and deterioration of the marble bed support plates. Detailed discussions of these problems have been previously published;^{4,5,6} however,

most of the problems have been resolved by cooperative efforts of the scrubber vendor and the NSP engineering staff. At present, NSP has converted one marble bed module to a spray tower in an on-going investigation to reduce maintenance costs.

MINNESOTA POWER AND LIGHT COMPANY (MP&L)

MP&L is currently constructing a 500-MW pc-fired boiler at the Clay Boswell Station at Cohasset, Minnesota. The boiler, called Clay Boswell Unit 4, will burn subbituminous coal from the Rosebud seam of the Big Sky mine at Colstrip, Montana, and has the following typical analysis: 24.5 pct moisture, 9.4 pct ash, 0.54 pct sulfur, and a HHV of 19,239 J/g (8,287 Btu/lb). An analysis of the coal ash is presented in Table 4.

Table 4. MP&L COAL ASH ANALYSIS

	Pct
Silicon dioxide, SiO ₂	43.3
Aluminum oxide, Al ₂ O ₃	17.0
Ferric oxide, Fe ₂ O ₃	5.7
Titanium oxide, TiO ₂	0.8
Phosphorous pentoxide, P ₂ O ₅	0.1
Calcium oxide, CaO.....	12.0
Magnesium oxide, MgO.....	6.7
Sodium oxide, Na ₂ O.....	0.2
Potassium oxide, K ₂ O.....	0.9
Sulfur trioxide, SO ₃	13.3

Clay Boswell Unit 4 will be required to meet a particulate emission standard of 0.04 g/MJ (0.1 lb/MM Btu), and a sulfur dioxide emission standard of 0.52 g/MJ (1.2 lb/MM Btu). A flue gas cleaning system, designed and constructed by Peabody Process Systems, will provide both particulate and sulfur dioxide control. A schematic of the FGD system is shown in Figure 3.

The flue gas cleaning system consists of four parallel modules; three modules are designed to handle flue gas from the boiler at full load, and the fourth module serves as a spare. Each module consists of one radial flow venturi for particulate removal, and one spray tower for sulfur dioxide removal. Reheat for the cleaned gas will be provided by mixing with flue gas taken upstream of the air preheater and cleaned of particulate by two electrostatic precipitators. The system is designed to operate using lime; however, it is planned to use the alkalinity present in the fly ash. If the fly ash present in the flue gas from Unit 4 is not sufficient to meet the sulfur dioxide emission limit, fly ash collected from baghouses on Units 1 and 2 will also be used. The unit will have a backup lime system available for use during periods of high levels of sulfur dioxide.

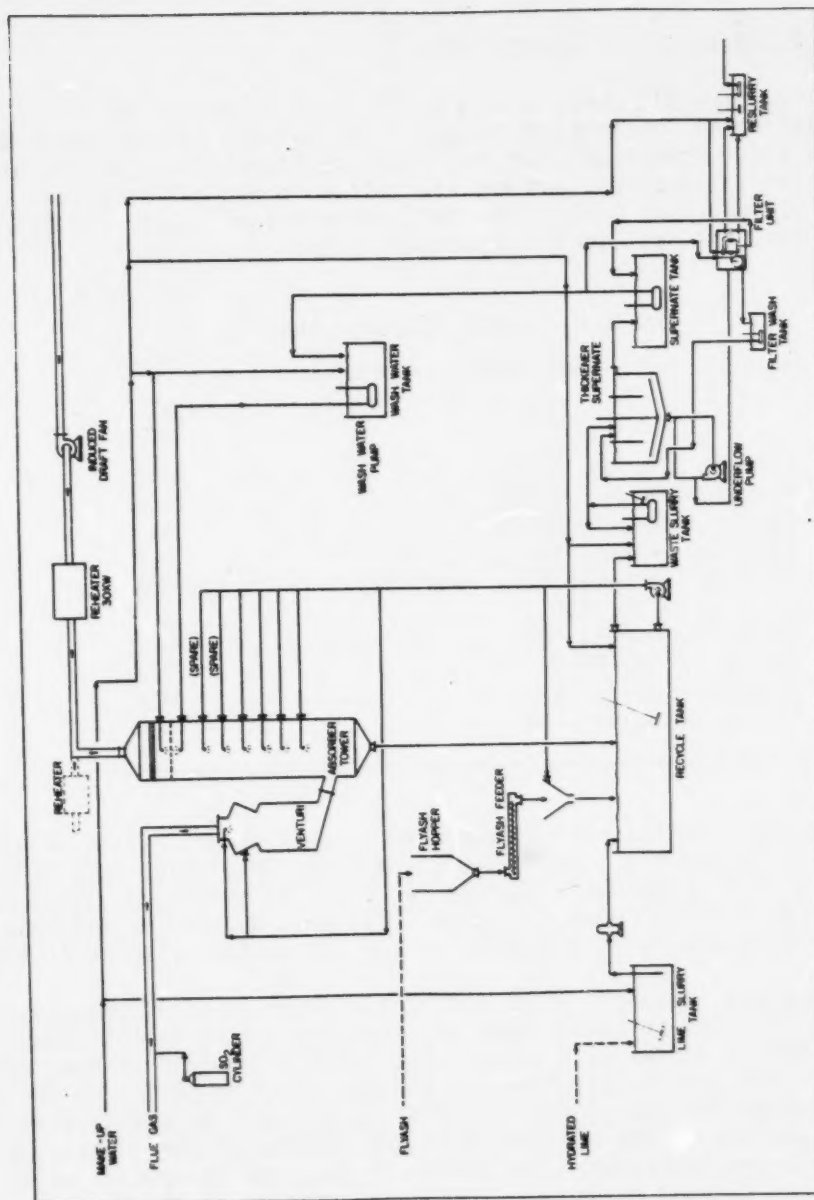


Figure 3. Flow schematic for Clay Boswell FGD pilot plant, Minnesota Power & Light Co.

Control of calcium sulfate scaling will be accomplished primarily by solids circulation and liquor retention in the slurry mix tank. The level of suspended solids in the scrubber solution will be maintained at about 12 pct to provide seed crystals for precipitation of calcium sulfate. The recirculated scrubbing liquor will have a 10-minute retention time in the reaction tank to allow desaturation of calcium sulfate. Pilot plant data indicate that the state of oxidation of the sulfur species is essentially 100 pct sulfate. Sludge solids will be disposed of in a pond.

The design and operating data were obtained on a 1-MW pilot plant wet scrubber using a sidestream of flue gas from Unit 3, which has a boiler similar in design to Unit 4 and uses the same coal. Tests at the pilot plant scrubber are continuing; however, preliminary results indicate that sufficient alkali can be dissolved from the fly ash to meet emission requirements without supplemental lime. No unusual chemistry-related operating problems have been encountered. Some selected results are presented in Table 5.

Table 5. SELECTED RESULTS OF MP&L PILOT PLANT PROGRAM

	Test Date: 4-19-78	5-3-78	5-4-78
SO ₂ :			
Inlet.....ppm..	1005	684	823
Outlet.....ppm..	430	219	305
Removal.....pct..	57.2	68	62.9
L/G.....	58	62	64
Fly ash utilization.....pct..	0	~100	~100
Slurry pH.....	5.4	3.2	3
Lime requirement, full-scale unit...metric ton/hr..	2.1	0	0

The L/G ratio represents the sum of the L/G values to the venturi and the spray tower. The results indicate that nearly 100 pct of the fly ash calcium can be utilized and that the sulfur dioxide removal efficiency is sufficient to meet the required emission standard. If similar results are obtained on the full-scale FGD unit, the lime reagent savings would be substantial.

MONTANA POWER COMPANY (MPC)

MPC operated a two-year pilot plant test program at the J. E. Corette Station in Billings, Montana. The pilot plant program was designed to investigate particulate removal and sulfur dioxide removal using the captured alkaline fly ash. Based on results of the pilot plant test program,

the scrubber vendor, Combustion Equipment Associates (CEA) constructed full-scale flue gas cleaning systems on two 360-MW units located at Colstrip, Montana. Colstrip Unit 1 was put into commercial operation in May 1976, and Unit 2 was put in commercial operation in August 1976. The information presented in this report was obtained from earlier publications^{7,8,9} and from Dr. Carlton Grimm, Montana Power Company.

The boilers burn coal from the Colstrip mine, which has 8.2 pct ash, 0.78 pct sulfur, and a HHV of about 20,082 J/g (8,650 Btu/lb). The composition of the fly ash is shown in Table 6.

Table 6. FLY ASH COMPOSITION AT THE COLSTRIP STATION, MONTANA POWER COMPANY*

	Pct
Silicon dioxide, SiO ₂	41.6
Aluminum oxide, Al ₂ O ₃	22.4
Titanium oxide, TiO ₂	0.8
Ferric oxide, Fe ₂ O ₃	5.4
Calcium oxide, CaO.....	21.9
Magnesium oxide, MgO.....	5.0
Sodium oxide, Na ₂ O.....	0.3
Potassium oxide, K ₂ O.....	0.1
Phosphorous pentoxide, P ₂ O ₅	0.4

* Sulfur trioxide-free basis.

The flue gas cleaning system consists of a downflow venturi located in an upflow spray tower contactor, as shown in Figure 4. The venturi is equipped with a variable throat to maintain a constant pressure drop at variable loads. Each flue gas cleaning system on the two units are comprised of three modules, with each module capable of handling 40 pct of full load. Slurry from the reaction tank is pumped to the venturi (L/G 15) for particulate control, and to the spray tower (L/G 18) for sulfur dioxide absorption. The flue gas leaving the spray section passes through a wash tray which is designed to remove entrained solids which could foul the mist eliminator. The saturated flue gas is then reheated using steam coils and discharged to the atmosphere.

The scrubber system is operated in a closed-loop mode. Recycle slurry from the absorber tower recycle tank is bled to an intermediate pond to settle the solids, and the clarified liquor returned to the scrubber system. Part of the pond return liquor is diluted with makeup water and used for bottom wash of the mist eliminator. The remaining portion of the pond water is used to wash the bottom of the wash tray. Makeup water is added to the pond return water to minimize calcium sulfate saturation levels in the mist eliminator washwater. To date there have not been any significant scaling problems associated with the demisters.

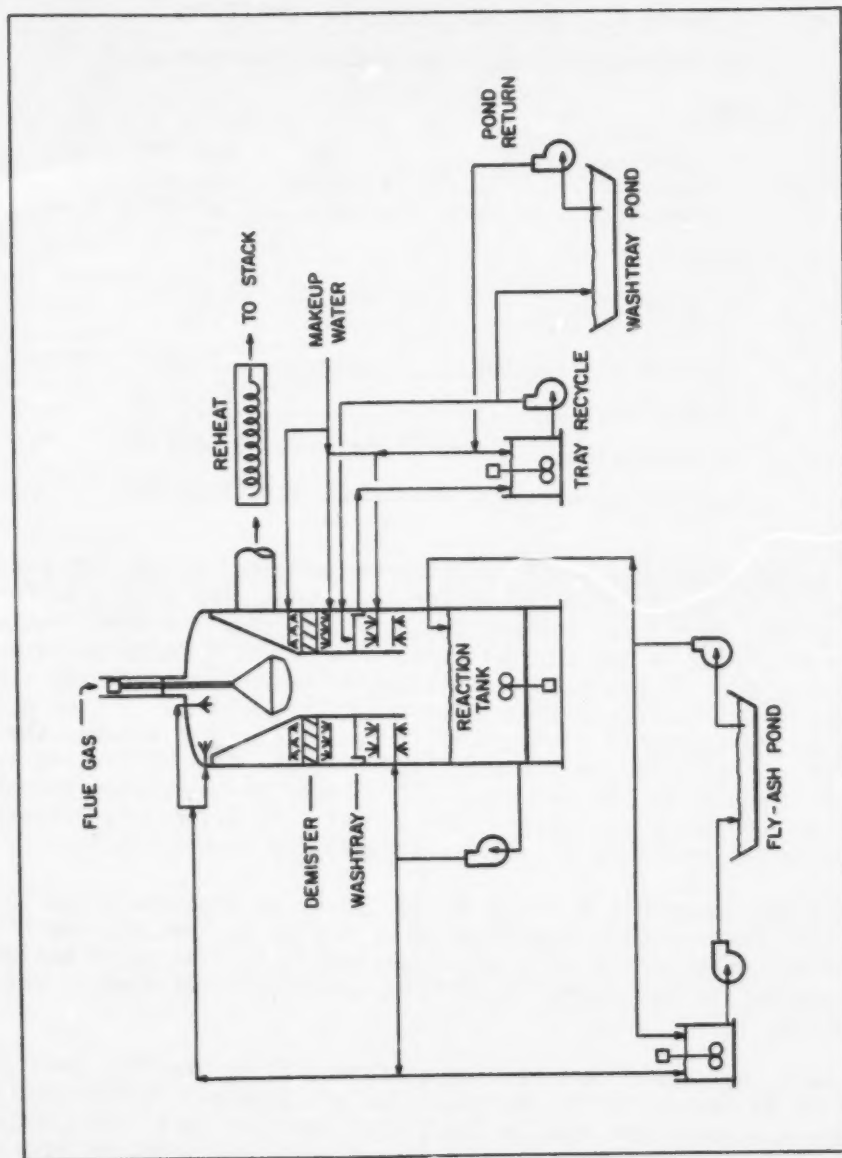


Figure 4. Flow schematic for Colstrip FGD system, Montana Power Company.

The scrubber system is operated so that maximum fly ash calcium is utilized while achieving the desired sulfur dioxide control and minimizing calcium sulfate scaling. A summary of the sulfur dioxide removal efficiencies and utilizations are presented in Table 7.

Table 7. MPC FGD OPERATING CONDITIONS

SO ₂ :	
Inlet.....ppm..	600-750
Outlet.....ppm..	50-200
Removal.....pct..	80-90
L/G:	
Venturi.....	15
Absorber tower.....	18
Fly ash utilization.....pct..	70-90
Slurry pH.....	4.5-5.5
Lime requirement..metric ton/hr..	0.19-0.27

The sulfur dioxide removal efficiencies obtained by the FGD system range from 80 to 90 pct when the boiler is burning coal with a sulfur content varying from about 0.6 pct to about 0.9 pct. The contribution of the fly ash calcium to the removal efficiency varies from 65 to 90 pct, and the corresponding approximate utilization values based on fly ash containing 20 pct calcium oxide ranges from 70 to 90 pct. Slaked quicklime is added to the system as a slurry if it is necessary to augment the fly ash alkali and maintain the scrubber liquor at the desired pH. At average coal sulfur conditions, the amount of lime added to the system represents about 10 pct of the total alkali. The annual lime savings is estimated to be 10,000 metric tons/year (11,000 tons/year) for each FGD unit.

Since the operating pH range of the slurry is moderately low to leach calcium from the fly ash, other species will also be leached, and high levels of magnesium and sulfate have been reported. The major sulfur species present in the slurry is sulfate, which presents over 90 pct of the total sulfur.

To date, no significant calcium sulfate scaling problems have occurred. The scaling is controlled by recirculation of suspended solids (about 12 pct), adequate retention time in the slurry reaction tank, and control of the slurry pH at a constant value. Other operational problems have been previously reported.^{7,8,9}

SQUARE BUTTE ELECTRIC COOPERATIVE (SBEC)

SBEC is currently operating a 238-MW (Unit 1) and a 450-MW (Unit 2) cyclone-fired generating unit at the Milton R. Young Station at Center, North Dakota. The primary fuel is a North Dakota lignite which has the following typical analysis: 35 pct moisture, 8 pct coal ash, 0.8 pct sulfur, and a HHV of 15,555 J/g (6,700 Btu/lb). An analysis of a typical fly ash is presented in Table 8.

Table 8. TYPICAL ANALYSIS OF LIGNITE FLY ASH PRODUCED BY CYCLONE-FIRED CENTER UNIT 1 AT THE MILTON R. YOUNG STATION

	Pct
Loss on ignition at 800° C.....	2.2
Silica, SiO ₂	29.8
Aluminum oxide, Al ₂ O ₃	12.7
Ferric oxide, Fe ₂ O ₃	10.6
Titanium oxide, TiO ₂	0.5
Phosphorous pentoxide, P ₂ O ₅	0.3
Calcium oxide, CaO.....	25.7
Magnesium oxide, MgO.....	4.5
Sodium oxide, Na ₂ O.....	2.2
Potassium oxide, K ₂ O.....	2.0
Sulfur trioxide, SO ₃	6.4

The 450-MW unit is required to meet the Federal emission standard of 0.52 g/MJ (1.2 lb SO₂/MM Btu). Particulate control for both units is provided by electrostatic precipitators, and sulfur dioxide control on Unit 2 is provided by a wet scrubber using fly ash alkali.

A pilot plant test program was conducted under a cooperative agreement by SBEC, Minnesota Power and Light Company, Sanderson and Porter Consulting Engineers, the Department of Energy's Grand Forks Energy Technology Center, and Combustion Equipment Associates-Arthur D. Little (CEA-ADL). The pilot plant results have been reported previously.^{10,11,12,13,14,15} System design and operating criteria were generated for construction and operation of a full-scale spray tower system on Unit 2 which would use fly ash alkali with supplemental lime.

A schematic of the full-scale scrubber system is shown in Figure 5. The system consists of two modules, with each module designed to handle 60 pct of the flue gas at full load. Each module contains a reaction tank, six spray nozzle banks, a wash tray, and demisters. Approximately 15 pct of the inlet flue gas bypasses each absorber tower and is used for reheat. The spray tower, reaction tanks, and mix tanks are coated with flakeglass lining for corrosion control; all pumps are rubber-lined for erosion/corrosion control. Solids are controlled by bleeding to a thickener. When the

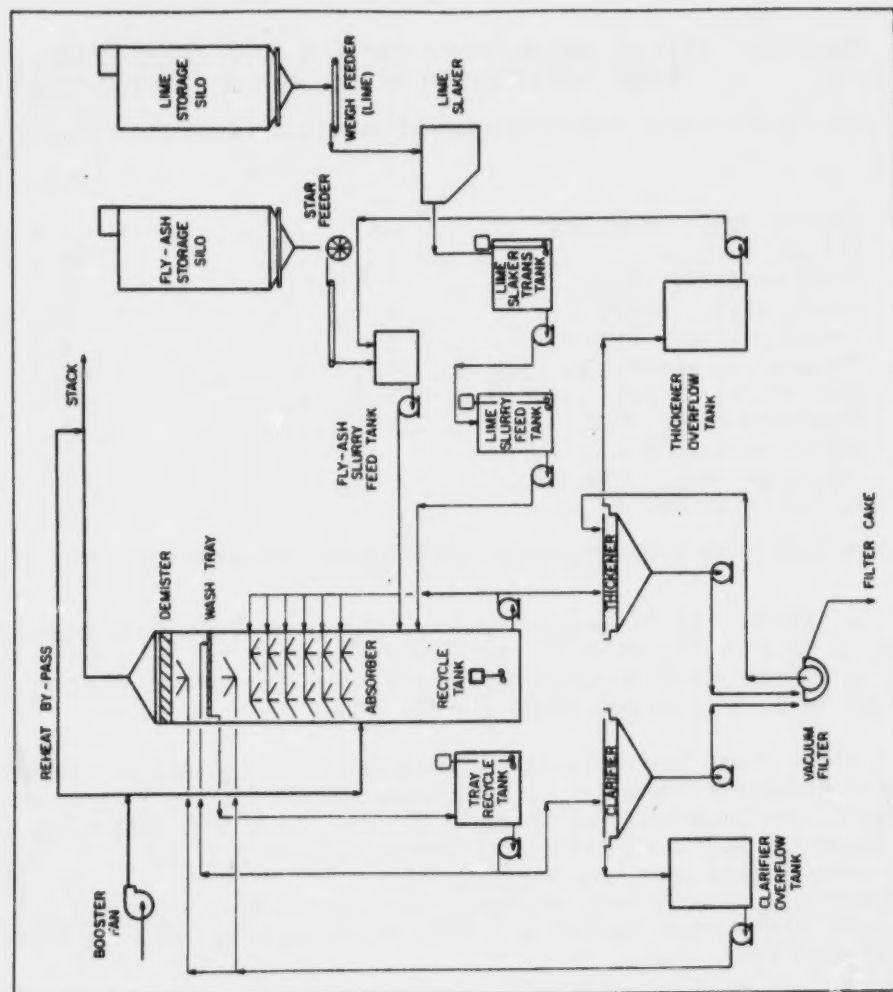


Figure 5. Flow schematic for Milton R. Young FGD system, Square Butte Electric Cooperative

thickener underflow contains 35 to 50 pct solids, they are pumped to a rotary vacuum filter and dewatered to approximately 60 to 80 pct solids. Dewatered sludge from the vacuum filter is disposed of in the strip mine. The state of oxidation of the sulfur species is essentially 100 pct sulfate.

The sulfur dioxide removal efficiency for flue gas passing through the spray towers when burning coal with average sulfur content has been 90 pct or better, using only fly ash alkali. Some selected results are presented in Table 9. The removal efficiency data represent only the spray tower efficiency; removal efficiency for the FGD system would be less since approximately 15 pct of the flue gas is bypassed and used for reheat.

Table 9. SBEC FGD OPERATING CONDITIONS

Absorber Tower SO ₂ :	
Inlet.....	ppm.. 450-900
Outlet.....	ppm.. 10-50
Removal.....	pct.. 90-99
L/G.....	80
Fly ash utilization.....	pct.. 75-100
Slurry pH.....	4-5
Lime requirement.....	metric ton/hr.. 0

Calcium sulfate scaling in the absorber tower and slurry lines is controlled by circulation of suspended solids, by sufficient liquor retention time in the recycle slurry tank, and maintaining the slurry at a pH range of 4 to 5. The recycle tank has a liquor retention time of about 8 minutes to allow calcium sulfate desupersaturation. The level of suspended solids in the recycle slurry is maintained at about 12 pct to provide seed crystals for precipitation of calcium sulfate. Scaling in the fly ash wetting tank has been a problem and requires cleaning every two to three months. Scaling has also occurred in the lime slurry feed line. Neither calcium sulfate scaling nor deposits have been observed on the wash tray or mist eliminator.

The recycle slurry is maintained at a pH range of 4.0 to 5.0 by the addition of only fly ash alkali. Fly ash from both Unit 1 and Unit 2 is available for use, and is stored in a single storage silo. Operation of the FGD system at this pH range minimizes significant problems in the thickener and leaches 75 to 100 pct of the calcium alkali from the fly ash. Pilot plant data indicate that significant amounts of other fly ash species such as sodium, magnesium, and trace elements, can also be solubilized, resulting in high levels of dissolved solids. The level of dissolved solids reached an equilibrium value of about 10 pct in the pilot plant operation. The level of dissolved solids accumulated during the limited operation of the full-scale system is only 2 to 3 pct.

The system availability has been poor, averaging only 46.1 pct. The scrubber system problems have been primarily mechanical. However, one problem relating to the chemistry of the fly ash alkali which is responsible for a significant amount of downtime is associated with the thickener tank and the slurry pH.

The slurry is maintained at a pH range of 4.0 to 5.0 by the addition of fly ash alkali. If the slurry rises above pH 5, the settled solids in the bottom of the thickener begin to exhibit pozzolanic activity, which eventually results in a hardened mass requiring shutdown of the thickener (and the FGD system) and manual removal of the sludge. Similar problems have occurred in the vacuum filter dump bin. Additionally, the thickener is operated so that the thickener underflow liquor is maintained in a pH range of 4.5 to 6.0. If the thickener underflow liquor pH rises above 7.0, the sludge material becomes slimy, resulting in poor dewatering and caking properties.

If the pH of the recycle slurry is too low, the suspended solids in the thickener overflow increases. Data generated during the pilot plant program indicated that the solids settling properties and filter characteristics degraded as the pH of recycle slurry decreased. Similar observations have been observed during operation of the full-scale system. SBEC is currently considering the use of a centrifuge for solids separation.

The poor settling characteristics of the suspended solids can be attributed to a decrease in the particle size distribution. Figure 6 illustrates the particle size distribution of sludge collected during the pilot plant program. In general, as the pH of the scrubber slurry decreases, the particle size of the recirculated suspended solids decreases. Preliminary results on the sludge particles using scanning electron microscope techniques (SEM) indicate that calcium sulfate crystals are not as large at low pH conditions as they are at high pH conditions. The SEM investigation also indicates that some calcium sulfate is precipitating on the surface of fly ash particles.

Additional mechanical or electrical problems resulting in significant downtime are due to the booster fans, liquid level monitors, flue gas bypass dampers and chains (freeze-up during cold weather operation), slurry spray pumps, erosive failure of flakeglass lining in one spray tower, and the fly ash feed system.

The problems encountered with the slurry spray pumps are caused by an accumulation of 1.25 cm (1/2-inch) or greater solids in a strainer immediately upstream of the pump. As the slurry flow became restricted, the rubber lining would be pulled from the pump casing, resulting in a significant number of plugged spray nozzles. At the present time, pressure drop across the strainers is monitored by the operator and manually cleaned when necessary. The solids, formed in the slurry recycle tank, are caused primarily by fluctuations in the liquor level. When the liquor level decreases, slurry containing suspended solids dried on the tank walls. As the deposit thickness increases due to continuous fluctuations in the level of recycle slurry liquor, the deposits eventually flake off and are caught in the strainers.

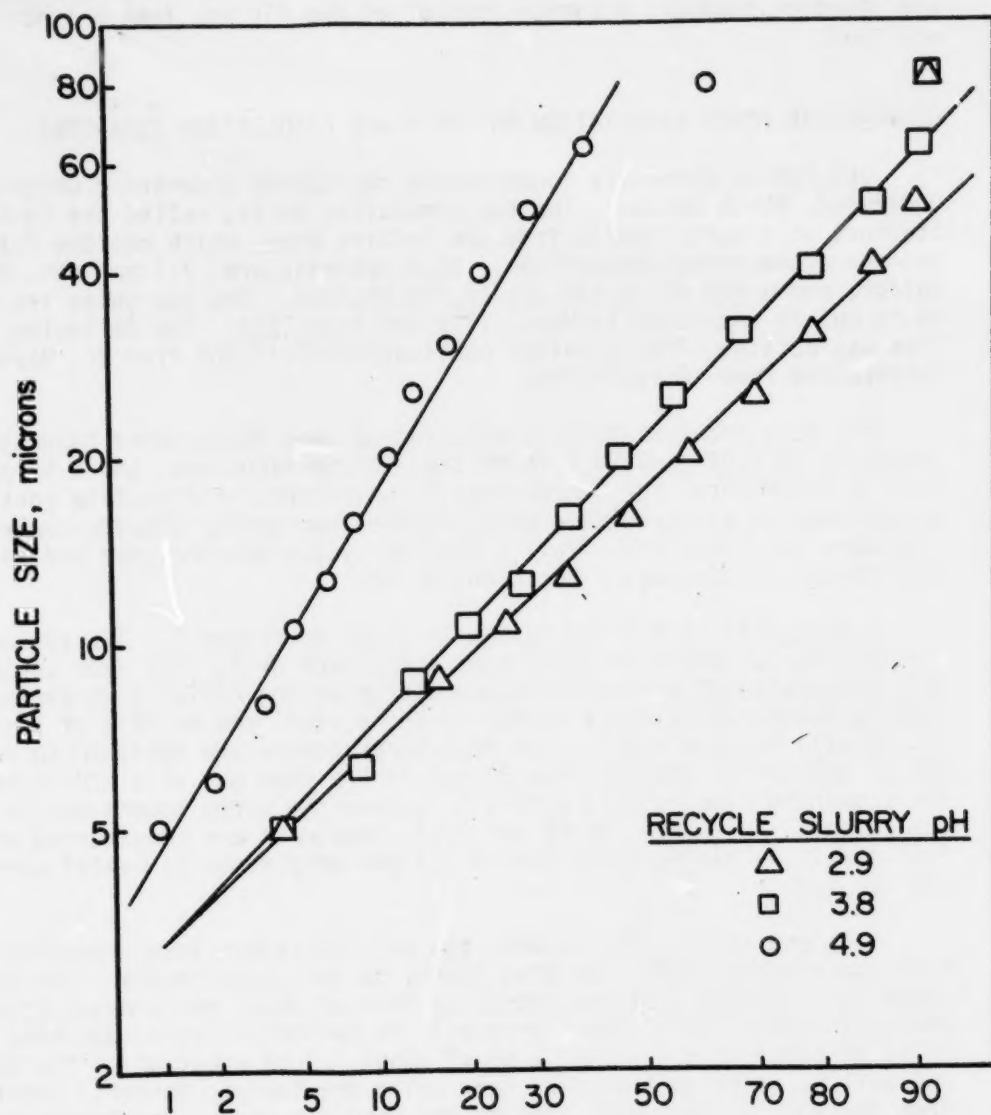


Figure 6. Particle size distribution of fly ash alkali sludge from Square Butte Electric Cooperative pilot plant wet scrubber.

The fly ash delivery system was designed to add fly ash to a wetting tank using a conveyor weigh belt. However, the delivery system was not adequate since the fly ash would flow off the sides of the conveyor belt (similar to water). At the present time, the belt has been replaced by a star feeder; however, accurate control of the fly ash feed has not yet been achieved.

COOPERATIVE POWER ASSOCIATION-UNITED POWER ASSOCIATION (UPA-CPA)

UPA-CPA is currently constructing two 550-MW generating units near Underwood, North Dakota. The two generating units, called the Coal Creek Station, will burn lignite from the Falkirk Mine, which has the following average as-received composition: 39.8 pct moisture, 7.1 pct ash, 0.63 pct sulfur, and a HHV of 14,530 J/g (6,258 Btu/lb). The two units are expected to be put in operation in March 1979 and late 1979. The following information was obtained from previous publications^{16,17} and from Mr. Wayne Hickok, Cooperative Power Association.

The Coal Creek Station is required to meet the current Federal standards of 0.04 g/MJ (0.1 lb/MM Btu) for particulates, and 0.52 g/MJ (1.2 lb SO₂/MM Btu) for sulfur dioxide emissions. Particulate control will be provided by electrostatic precipitators and sulfur dioxide control by wet scrubbers using fly ash alkali. The FGD system was designed and is being constructed by Combustion Engineering, Inc.

A schematic of the FGD system is shown in Figure 7. The system will contain four pressurized spray towers for each unit. The four spray towers will be capable of operating independently as the boiler load varies. When lignite containing average sulfur is being used, one or more of the spray towers will be on standby. The four spray towers are designed to remove 90 pct of the sulfur dioxide from 60 pct of the flue gas at a L/G ratio of 60. The remaining flue gas (40 pct) will bypass the spray towers and be used for reheat. The spray towers and outlet ductwork are constructed of stainless steel containing a minimum of 2.5 pct molybdenum to resist corrosion and erosion.

Each FGD system will contain two unlined carbon steel reaction tanks, with one reaction tank providing slurry to two spray towers. The reaction tanks will provide a slurry retention time of about ten minutes (two spray towers in operation). The slurry will be maintained at a suspended solids level of about 15 pct and at a pH of about 7.0 by addition of fly ash. Fly ash will be added to a wetting tank using gravimetric feeders; overflow from the wetting tank will feed to the reaction tank. Supplemental lime can be added when required. UPA-CPA calculations indicate that capital costs for equipment, ash silos, and building space for the fly ash system can be recovered in about four years of operation if the fly ash alkali reduces the lime requirements by 50 pct.

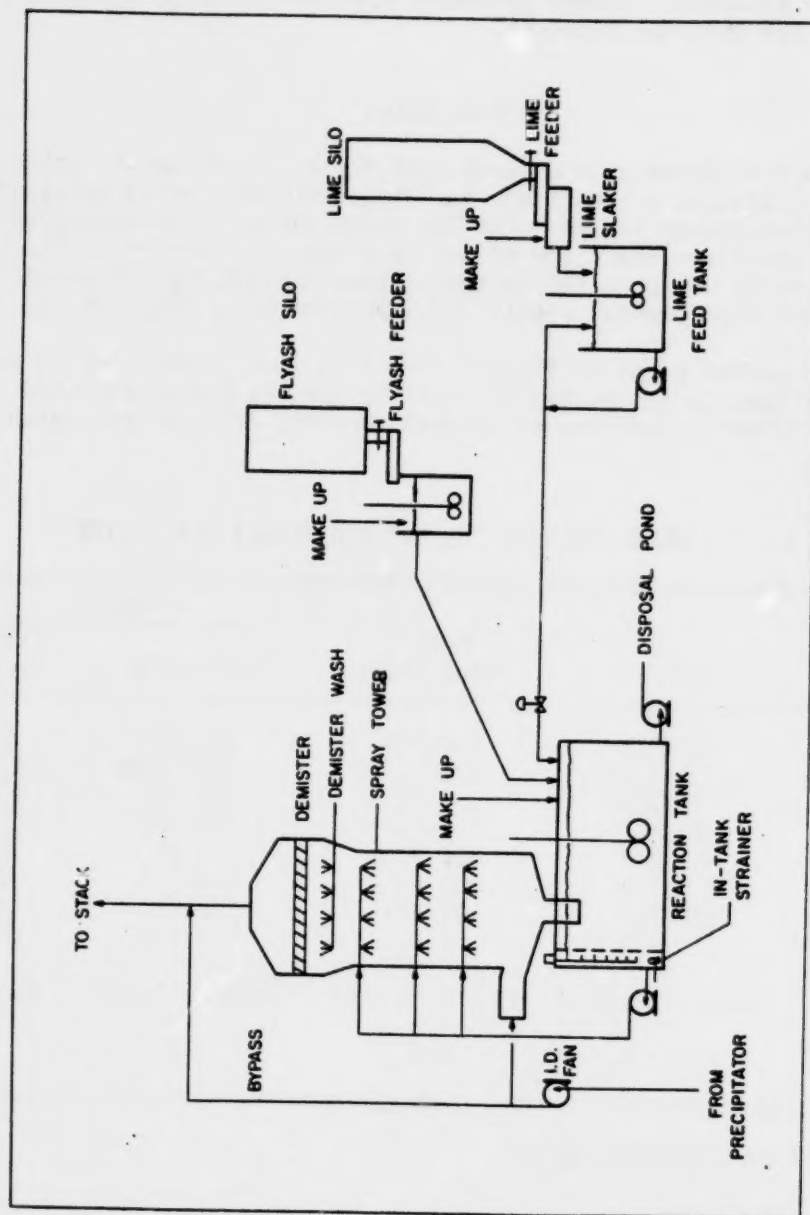


Figure 7. Flow schematic for Coal Creek FGD system, United Power Association-Cooperative Power Association.

Scaling in the FGD system will be controlled by maintaining the suspended solids concentration at about 15 pct, providing adequate slurry retention time in the reaction tank (about ten minutes), and maintaining the pH at a constant value (about pH 7). Makeup water will be provided by cooling water blowdown. Sludge generated from the scrubber system will be pumped to an ash pond for disposal.

DISCUSSION

If the alkali present in Western coal fly ash is to be utilized, in part or completely, as a substitute for lime/limestone, it is necessary to operate at a moderately low pH. The FGD units currently in operation vary in operating pH value from a low of 4.0 to a high of about 5.5. The actual pH operating range for each FGD system varies, depending on site-specific factors and the supplemental alkali (lime or limestone) added to the system.

One consequence of using fly ash alkali at a moderately low pH value is that other species in the fly ash will be leached and accumulated in the scrubber liquor. Analyses of several scrubber liquors are presented in Table 10.

Table 10. LIQUOR ANALYSES FOR FLY ASH ALKALI FGD SYSTEM*

	NSP	MP&L Pilot Plant	SBEC	
			Full-scale	Pilot Plant
Calcium	520-570	630-670	430-550	530
Magnesium	2,700-3,900	635-1,300	500-1,600	13,000
Sodium	275-350	5-15	400-880	13,000
Potassium	40-65	1.5-24	130-500	1,010
Aluminum	70-275	180-380	1.3-42	329
Iron	1.8-5.9	25-63	0.5-2.3	808
Chloride	480-765	72-127	72-370	200
Fluoride	2.3-6.0	<1.0	2.7-7.3	<1.0
Sulfite	0	0	0	0
Sulfate	14,500-36,500	12,000-15,700	13,300-19,000	65,500
Dissolved solids	2.2-4.6	1.8-2.3	2-4	10

* Concentration units are mg/l.

The data presented in Table 10 were generated using a series of liquor samples collected during conditions believed to be representative of the NSP and MP&L plants. The data presented for the SBEC full-scale scrubber are not considered to be representative of steady-state operation since the samples were collected after only about 30 days of continuous operation.

and pilot plant data indicate that two to three months are required for ionic equilibrium. For comparative purposes, an analysis of the pilot plant liquor is also included in Table 10. The pilot plant liquor was collected after about six months of operation, and the dissolved solids had reached a steady-state level of about 10 pct. As can be seen, the levels of magnesium, sodium, and sulfate are present in appreciable concentrations; similar concentrations are expected in the full-scale unit, but will be dependent on the degree of closed-loop operation.

The high values of dissolved solids are beneficial to the ash alkali chemistry since the system can be operated at a lower slurry pH compared to open-loop operation, and still maintain efficient sulfur dioxide removal. Recent results at the MP&L plant plant¹⁸ indicate that a 63 pct sulfur dioxide removal efficiency can be achieved during closed-loop operation (L/G 62, high dissolved solids) at a slurry pH of about 3.0. Similar tests conducted during open-loop operation (L/G 62, low dissolved solids) indicate that a pH in the range of 4.8 to 5.0 would be required to obtain a similar sulfur dioxide removal.

The sludge generated from the fly ash FGD systems generally contain significant amounts of sulfate as compared to sulfite. The SBEC FGD unit and the MP&L pilot plant wet scrubber contain essentially 100 pct sulfate in the sludge waste. The sludge produced at NSP contains essentially all sulfate, since air oxidation is employed. Without forced air oxidation, the scrubber liquid would contain about 25 to 40 pct of the sulfur species as sulfite, and the physical properties of the sludge would be modified. Some physical characteristics of the various ash alkali sludges are presented in Table 11.

Table 11. PHYSICAL CHARACTERISTICS OF ASH ALKALI FGD SLUDGE

	SBEC	MP&L	NSP
Moisture.....pct..	21.6	17.1	19.7
Permeability.....cm/sec..	$0.1-4 \times 10^{-5}$	5×10^{-5}	2.8×10^{-5}
Unconfined compressive strength.....kg/cm ² ..	1.27	0.81	1.4
Secant modulus of elasticity.....kg/cm ² ..	39.1	19.4	22.2
Surface area.....m ² /g..	19.0	7.36	7.05

The compressive strength and modulus of elasticity values were obtained from the stress-strain profiles shown in Figure 8. The profiles are indicative of the unreacted fly ash alkali in the sludge. In general, the MP&L sludge had the highest calcium alkali utilization and exhibits the lowest stress value. The NSP sludge had the lowest calcium alkali utilization,

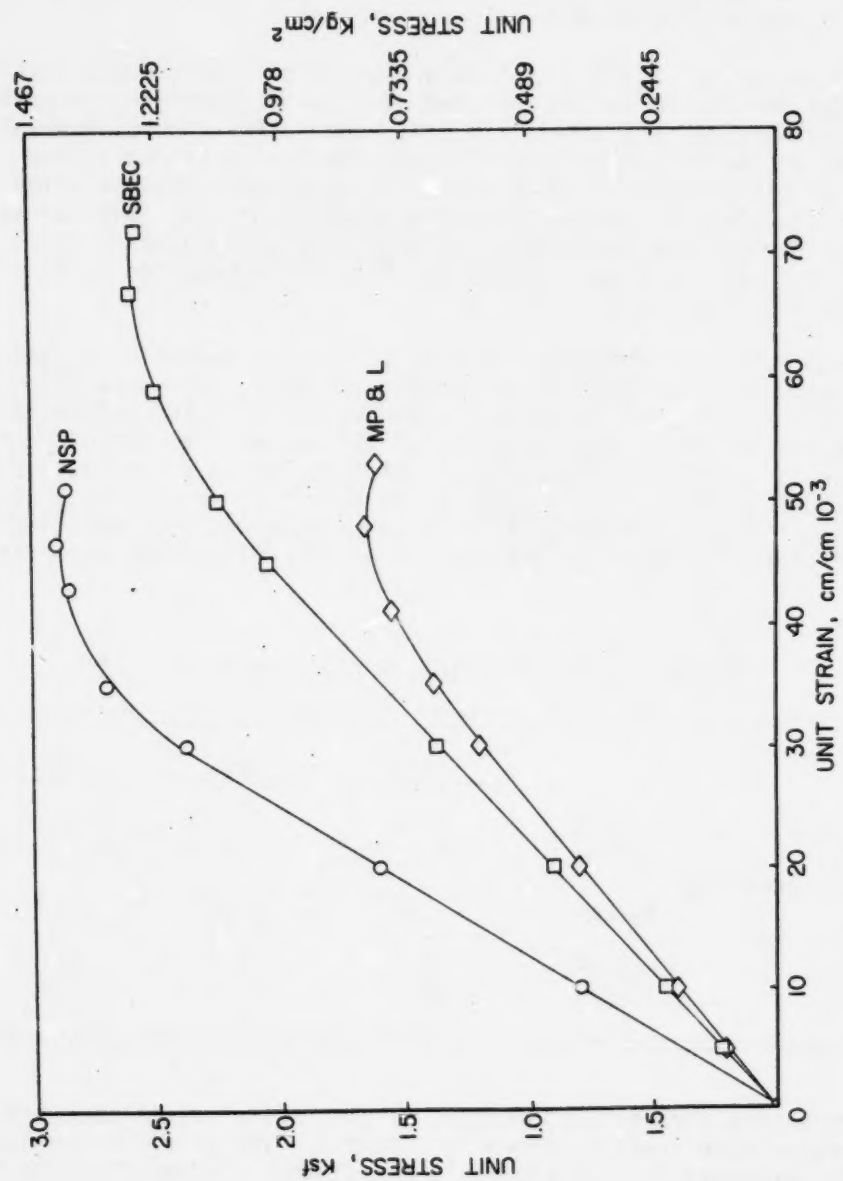


Figure 8. Stress-strain profiles for three fly ash alkali sludges.

in addition to a reported low limestone utilization, and exhibits the highest compressive strength. In general, however, all three sludges are structurally suitable for landfill disposal without additional fixation agents. At present, the SBEC FGD sludge is being disposed of in the stripmine, and studies on structural stability and potential groundwater contamination are being conducted by the University of North Dakota under EPA funding.

In conclusion, the utilization of Western alkaline fly ash, either partially or completely, as a substitute for lime or limestone in a wet scrubber FGD system has been successful. The present report does not attempt to detail economic savings. Actual savings will be site-specific, depending on plant location, scrubber system design, and mode of operation; however, substantial savings can be realized because of reduced lime or limestone requirements.

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Environmental Effects of FGD Disposal
A Laboratory/Field Landfill Demonstration

by

N. C. Mohn, A. L. Plumley, A. L. Tyler
Kreisinger Development Laboratory
Combustion Engineering, Inc.
Windsor, Connecticut

and

R. P. Van Ness
Louisville Gas & Electric
Louisville, Kentucky

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ABSTRACT

Results of an EPA-funded demonstration program for the conversion of fly ash and chemically treated sludge from sulfur removal scrubbers (AQCS) into a stabilized landfill having structural or recreational applications are discussed. An earlier paper (TIS-5485) described the laboratory studies and the preparation of field sites for the disposal of this waste. This paper covers efforts by Combustion Engineering, Inc. and Louisville Gas & Electric Co. to monitor the disposal sites over a two-year period.

Properly prepared landfill from FGD sludge/fly ash mixtures will not contaminate the surrounding groundwater. Results obtained from analysis of leachates from the series of landfill impoundments in this study show that trace elements on the RCRA list of contaminants were found in concentrations below those established to characterize hazardous or toxic waste.

ENVIRONMENTAL EFFECTS OF FLUE GAS DESULFURIZATION WASTE DISPOSAL A LABORATORY/FIELD LANDFILL DEMONSTRATION

INTRODUCTION

The most extensive commercial experience in flue gas desulfurization has been with lime/limestone wet-scrubbers, and it is anticipated that they will account for the majority of sulfur or SO_2 removal systems at electric power stations for the next 10 to 15 years. One of the major challenges associated with the commercial development of these systems is the large amount of by-product sludge that must be disposed of within the constraints of land- and water-quality regulations. It has been estimated that within five years, air quality control regulations will require the desulfurization of flue gas from 60 million kW of electricity generating capacity per year. If this estimate is realized, over 30 million tons of ash-free by-product sludge (50% solids) will be produced per year.⁽¹⁾

Combustion Engineering has been working on means of disposal and possible uses of the by-products from flue gas desulfurization (FGD) systems since 1969 with the primary objective of providing environmentally safe methods for the disposal and/or utilization of both the solid and liquid wastes from lime/limestone FGD systems. This work has continued with a major laboratory/field demonstration in cooperation with Louisville Gas & Electric Co. of the techniques for landfill disposal of by-product sludges. The work was performed under an EPA contract with the Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina. This paper gives the results of the demonstration.

HISTORICAL BACKGROUND

During the past ten years, nearly fifty different procedures for direct disposal and utilization with varying degrees of beneficiation have been investigated. The technically more promising of these sludge utilization processes are indicated in Table 1. Previous papers have discussed these potential uses and special disposal methods.^(2,3) Based on our studies, however, we have concluded that utilization will not be able to eliminate the problem of disposal any more than the utilization of fly ash (10-15 percent of the annual production) has solved the problem of fly ash disposal. Consequently, most of the waste by-products from flue gas desulfurization will be disposed of in ponds or used as landfill. The choice of disposal methods and amount of treatment required will depend on the geographical requirements, economic considerations and the particular preferences of the operating company.

Prior laboratory work by Combustion Engineering and others had indicated the environmental advantages of the disposal of stabilized FGD by-product sludges over untreated sludges. Haas and Ladd⁽⁴⁾ showed that waste solids from a limestone scrubbing system could be stabilized by dewatering and subsequent mixing with clay soil or a Western type fly ash having a high alkali content. Further studies^(5,6) showed that the addition of fly ash

Table 1 POTENTIAL USES FOR FGD BY-PRODUCT SLUDGE

Major	1. Cement manufacture
	2. Concrete admixture for construction
	3. Fill material for land recovery
	4. Manufacture of sinter bricks
	5. Manufacture of gypsum board
	6. Manufacture of wall panels
	7. Manufacture of light weight aggregate
	8. Manufacture of mineral aggregate
	9. Production of mineral wool
	10. Recovery of calcium oxide
	11. Recovery of sulfur
	12. Road base in highway construction
	13. Soil stabilizer for embankment and water retaining structures
Other	14. Airport pavement mixture
	15. Asphaltic filter and wear-course aggregate
	16. Grouting agent in wells
	17. Filler for glass
	18. Filler for fertilizer
	19. Filler for paint
	20. Filler for plastic
	21. Filler for rubber
	22. Fill material in abandoned mines, for fire control
	23. Fill material in abandoned mines, for subsidence control
	24. Neutralization of acid mine drainage
	25. Manufacture of autoclaved bricks
	26. Manufacture of porous pipes
	27. Production of cenospheres for lightened building materials
	28. Reclamation of polluted lakes
	29. Recovery of aluminum
	30. Recovery of magnesium oxide
	31. Sand blasting grit
	32. Soil amendment
	33. Filter aid for sewage sludge dewatering

and/or lime to FGD solids resulted in the formation of a number of mineral compounds of high strength and low permeability.

Since the completion of the earlier work, it has become apparent that FGD sludges contain varying concentrations of trace elements and dissolved salts which, in an unstabilized state, could contaminate surface or ground water. Although some soils will absorb many of the trace elements in FGD sludge, major ions, e.g., calcium, sulfate and chloride may not be readily absorbed. Therefore, the impoundment of sludge must be accompanied by various stabilization procedures that allow a sufficient margin of safety to control or prevent seepage. Consequently, leachate analyses were added to the unconfined compressive strength tests and permeability studies that were already a part of the sludge-landfill stabilization program.

LG&E/EPA WASTE DISPOSAL PROGRAM

The laboratory/field landfill demonstration was designed to show the feasibility of producing structural landfill from mixtures of by-product sludge and fly ash. (Last year Louisville Gas & Electric Co. and Combustion Engineering, Inc. reported on the progress of this demonstration. Details of the program setup and procedures can be found in Ref. 7.) To be considered acceptable, a landfill must provide a material of sufficient structural integrity to meet minimum standards of a compressive strength $>1 \text{ ton/ft}^2$ (0.1 MPa) and permeability $<5 \times 10^{-5} \text{ cm/s}$.⁽⁸⁾ In addition, a landfill must not contaminate groundwater by leachate or surface water by runoff or erosion. The standards used were the levels established for defining leachates from hazardous wastes under Resource Conservation and Recovery Act (RCRA), section 250.13(d).⁽⁹⁾

The program was divided into two phases: laboratory and field demonstration. Although laboratory tests were run for the previous sludge-landfill studies,^(4,6) new laboratory tests were necessary to determine the optimum composition of materials and additives for the field tests, because the sludge for this demonstration had a somewhat different composition and the landfill would be deposited in a part of the country with a different climate from the sludge in the earlier studies. These tests provided the baseline values for the strength, permeability, and leachate quality of each mixture evaluated. The field demonstration was designed to provide similar information on the behavior of stabilized materials under natural environmental conditions including precipitation, and freeze/thaw. Included in the field phase was evaluation of the handling, transportation, and placement of the various sludge mixtures.

The by-products used in this demonstration were obtained from the wet scrubbing of flue gas from combustion of 3% sulfur West Kentucky coal at the 65 Mw steam generator (No. 6) at the Paddy's Run Station of LG&E. The FGD system at Paddy's Run was placed in operation in 1973 and has used carbide lime, a locally available by-product from an acetylene manufacturer as the absorbent. Phase I of the waste disposal program was designed to provide a demonstration of impoundment of mixtures of fly ash and chemically treated sludge from carbide lime scrubbing.

From the standpoint of general usage, commercial lime is more likely to be used as the SO_2 absorbent. Phase II of this program was conducted, therefore, using sludge obtained from scrubber operation with a commercial lime absorbent.

The flue gas desulfurization system consists of two scrubber modules which operate in parallel at full load. Figure 1 shows the overall arrangement of the scrubbing system during the collection of the by-product used during this study. Inlet SO_2 concentrations were about 2000 ppm at a gas flow rate of 180,000 Nm^3/hr (175,000 ACFM), with the boiler at half load. A liquid/gas ratio (L/G) of 7.5 L/Nm^3 (28 gal/1000 CFM) was maintained during the test program. For Phase I, Carbide Lime Operation, SO_2 removal ranged between 75 and 83 percent. A slurry inlet pH of 8 was controlled over the six week period required to collect the process sufficient by-product to fill six impoundments.

During Phase II, Commercial Lime Operation, about 2000 ppm magnesium was added to allow assessment of its effect on system operation. The slurry inlet pH of 8 was maintained and SO_2 removal exceeded 90%. The sludge by-produce was processed and all ten remaining impoundments were filled within one month.

Laboratory Tests

The ability to stabilize waste solids from AQC systems is a strong function of the moisture content of the solids. As the solids content of a sludge increases, the void ratio decreases producing a material with a higher dry density. Smaller quantities of stabilizing additives are required to harden sludges with low void ratios because the particles are in closer proximity to react with each other and the hardening agent. In addition, drier mixtures that are close to their optimum moisture content for compaction can be placed in the landfill at much higher densities than wet sludges.

The laboratory program evaluated a series of stabilized sludges, with different degrees of dewatering. Process I involved mixing thickener underflow with fly ash and a fixative to attempt to form a pumpable self-hardening mixture. Process II mixed a partially dewatered sludge with fly ash and a fixative to form a stable, compactable landfill. Process III combined a more highly dewatered sludge with fixative and/or fly ash.

Sludges from scrubbing processes using both carbide and commercial lime additives were used in the laboratory test. Sixty-two mixtures were prepared for the laboratory screening. The sludges were mixed with fly ash in ratios ranging from 0:1 to 1.5:1 parts by weight fly ash to dry scrubber solids. Varying percentages of fixative (lime, hydrated lime, carbide lime, or Portland cement) were added to aid the (cementing) reaction.

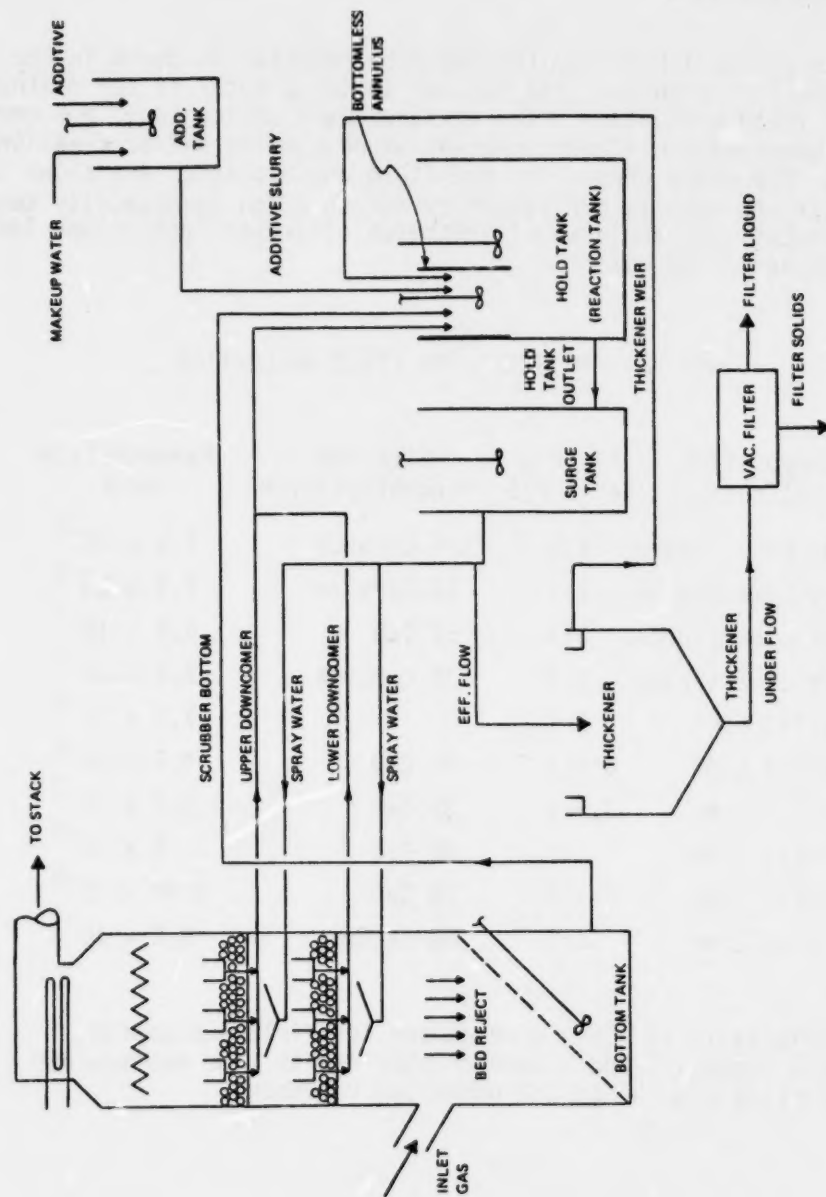


Figure 1 Flow arrangement for carbide and commercial lime testing

To predict the landfill behavior of stabilized sludges, the following series of tests were performed:

1. Unconfined compressive strength
2. Permeability
3. Leachate analysis

The results of the laboratory testing were reported in depth in the previous publication⁽⁷⁾. Briefly, from the lab tests, a total of ten mixtures were chosen for field evaluation. The mixtures were chosen to give a comparison between sludges with different degrees of dewatering and/or fixation additives. The mixes chosen for the field demonstration are shown in Table 2 with the results of laboratory strength and permeability tests. Table 3 presents the analyses of leachates collected from column leaching tests of three of the mixes.

Table 2 MIXTURES FOR FIELD EVALUATION

Field Mix No.	Sludge Composition Percent Solids/Type	Fly Ash Ratio F/S	Fixative Percentage/Type	Permeability cm/s	60-Day Compressive Strength Tons/ft ²
1	24% Carbide Lime Sludge	1:1	5% Carbide	7.6×10^{-5}	too soft
2	42% Carbide Lime Sludge	1:1	5% Carbide	2.9×10^{-6}	8.2
4	55% Carbide Lime Sludge	0:1	5% CaO	4.5×10^{-7}	9.2
6	55% Carbide Lime Sludge	1:1	3% Carbide	2.1×10^{-7}	25.1
7	65% Commercial Lime	1:1	0	7.0×10^{-6}	3.0
8	50% Commercial Lime	0.5:1	3% CaO	4.1×10^{-6}	7.3
9	50% Commercial Lime	1.5:1	3% CaO	5.7×10^{-7}	12.5
10	50% Commercial Lime	1:1	3% P.C.	5×10^{-5}	5.4
11	50% Commercial Lime	1:1	3% CaO	2.94×10^{-6}	8.5
12	50% Commercial Lime	1:1	3% Ca(OH) ₂	9.2×10^{-6}	7.1

Note: Originally 12 mixes were chosen for field evaluation. Mix numbers 3 and 5 were omitted due to time and weather constraints, hence the numbering sequence.

A quantity of the selected mixtures was prepared in the field with a process train designed for this purpose. The sludge/fly ash/treated mixtures were impounded in specially prepared sites designed to allow determination of leachate quality.

Table 3 LEACHATE ANALYSES

Pore Volume No.	Mix 1 24% Carbid Lime 1:1 Fly Ash Sludge 5% Carbid Lime		Mix 9 50% Comm. Lime 1.5:1 Fly Ash to Sludge 3% CaO		Mix 7 65% Comm. Lime 1:1 Fly Ash to Sludge 0	
	1 & 2	5 & 6	1 & 2	5 & 6*	1 & 2	5 & 6
Conductivity (S/cm ²)	3550	2350	1800	-	2850	2000
pH	11.4	11.1	9.3	-	7.8	7.7
TDS (ppm)	-	-	1100	-	1700	1200
Cl ⁻ (ppm)	115	<5	15	-	10	15
SO ₃ ⁼ (ppm)	-	65	230	-	40	-
Cd (ppm)	<0.01	<0.01	0.01	-	0.02	<0.01
Cu (ppm)	0.02	0.02	0.02	-	0.02	0.02
Pb (ppm)	0.1	0.1	0.1	-	0.1	0.1
Hg (ppm)	0.003	0.002	0.001	-	0.018	0.001
As (ppm)	-	0.03	0.01	-	0.03	0.03
SO ₄ ⁼ (ppm)	1580	1470	440	-	1580	1870
Ca (ppm)	300	310	6.7	-	320	300
Mg (ppm)	0.05	0.05	0.02	-	0.18	0.16
Se (ppm)	0.019	0.006	0.008	-	-	0.003

*Due to low permeability of samples, pore volumes 5 & 6 were not available after 60 days of collection.

Field Demonstration

A series of ten commercial above ground swimming pools and five larger impoundments (Figure 2) were used as monitored disposal sites for the sludge-fly ash mixtures.

Small Scale Impoundments

One type of disposal site consisted of small scale impoundments (Figure 3). The primary purpose of these test sites was to provide a means of determining the quality of the leachate and runoff from the test mixtures under field conditions. The impoundments were lined, above ground and have a capacity of about 25 cubic yards. A total of ten commercial above-ground swimming pools have been used as the small scale impoundments. Four of these were used for sludge mixtures made from carbide lime and six for mixtures using commercial lime.

The bottom six inches of each pool contained non-reactive graded gravel where the leachate, which permeated the test mixtures, was collected. The leachate was drained to a collection tank. The amount of leachate collected was estimated and, together with the National Weather Service Rainfall Data, was used to determine the average rate of leachate generation. The leachate from the impoundments was analyzed for dissolved ions. The first leachate samples were collected for analysis one week after filling each impoundment. Thereafter the leachate was collected and analyzed at two or three month intervals.

The runoff was collected from the small scale impoundment areas through a gravel filter held in place by a coarse screen. This procedure insured that drainage was always provided regardless of the level to which the sludge consolidated. The runoff was analyzed for dissolved species at the same intervals as that indicated for the leachate.

Large Scale Impoundments

The small scale impoundments provided a convenient means of determining maximum leaching rates and leachate quality without any interference from local surroundings. In the actual field site, the landfill will either absorb or release moisture to the surrounding soil. The large scale impoundment areas provided a means of assessing the impact of the disposal material in terms of its effect on local soil moisture and the quality (dissolved ions) of the moisture in the soil and of the water in the aquifer beneath the disposal sites.

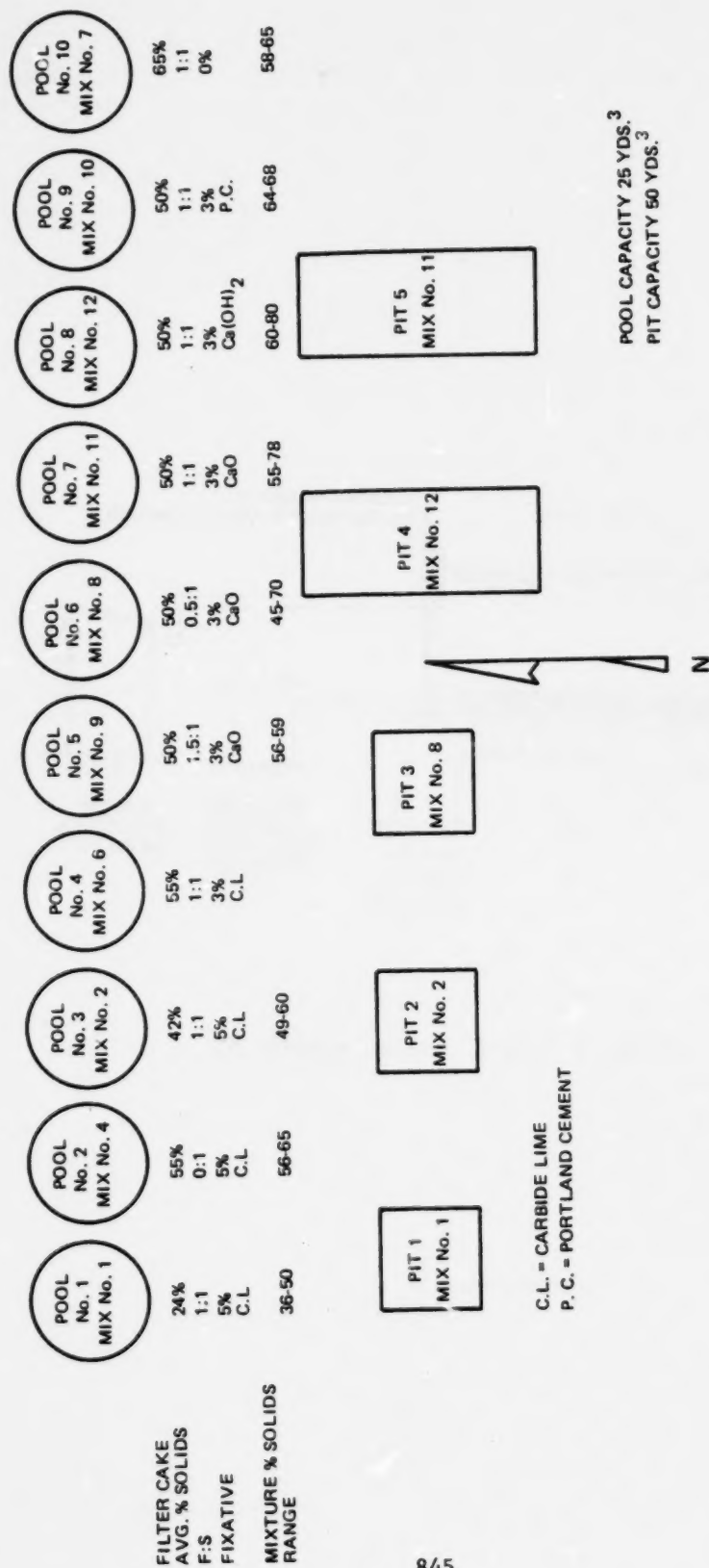


Figure 2 Sludge impoundment sites

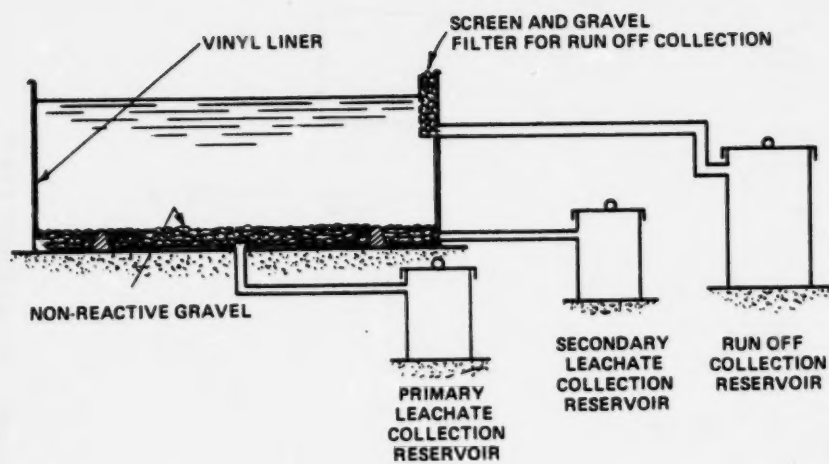


Figure 3 Small scale impoundment

Five large scale impoundment areas were excavated, each with a capacity of about 50 cubic yards. The disposal sites located in natural soil are of two styles: 10 foot x 10 foot x 8 foot deep pits and 30 foot x 8 foot x 4 foot deep pits. Two contained carbide sludge mixtures with the remainder containing mixtures of commercial lime sludge (Figure 4). Soil moisture was monitored by suction lysimeters located 6, 24 and 72 inches beneath the bottom of the test site. Particular attention was paid to the water collected in the lysimeters to determine if the soil is absorbing any dissolved ions from the sludge leachate. Leachate and ground water were analyzed periodically.

A schematic flow diagram of the waste material handling system used to process the sludge during the field demonstration phase is shown in Figure 5.

The entire thickener underflow was pumped around an 800 ft. circulation loop. A slip stream taken from the loop was used to fill a 10 ft. dia. x 10 ft. high slurry surge tank. The remaining slurry was then returned to the vacuum filter which is normally used to dewater the solids prior to disposal. Three separate processes were used to prepare the mixtures for disposal:

- For Process I, the sludge was pumped from the surge tank directly to the mixer into which additive and fly ash was being metered.
- In Process II, the sludge was dewatered in the filter press to produce a filter cake of the same solids content as the filter cake from the commercial rotary vacuum filter. When removed from the filter press, the filter cake fell into a surge bin from which it was metered into the mixer by means of a 6-inch variable speed screw (VSS) conveyor. Fly ash and additive were simultaneously metered into the mixer by a 4-inch VSS conveyor.
- Process III was planned to evaluate the use of a filter press operating at high pressure to dewater the sludge. The filter press provides a means of obtaining a much drier cake than can be obtained with a vacuum filter. A drier filter cake requires less fly ash and additive for stabilization. A 65% solids filter cake was expected, however, the maximum cake solids produced was 50%. The filter cake was stabilized with fly ash. All mixtures were discharged into trucks, transported seven miles and then placed as landfill.

Leachate Chemical Analyses

Due to freezing weather shortly after placement of the first series of lime mixes, very few liquid samples were collected from the small impoundments until the spring thaw, approximately four months later. Analyses of total dissolved solids (TDS) and selected trace elements in samples obtained from the inner ring or primary leachate reservoir (See Figure 3) and the overflow or runoff reservoir are depicted in Figure 6. The small impoundments represent the worst case for leachate generation, no attenuation by local soil is provided nor is any vegetation cover provided to minimize runoff. In addition, precipitation is held on the impoundments.

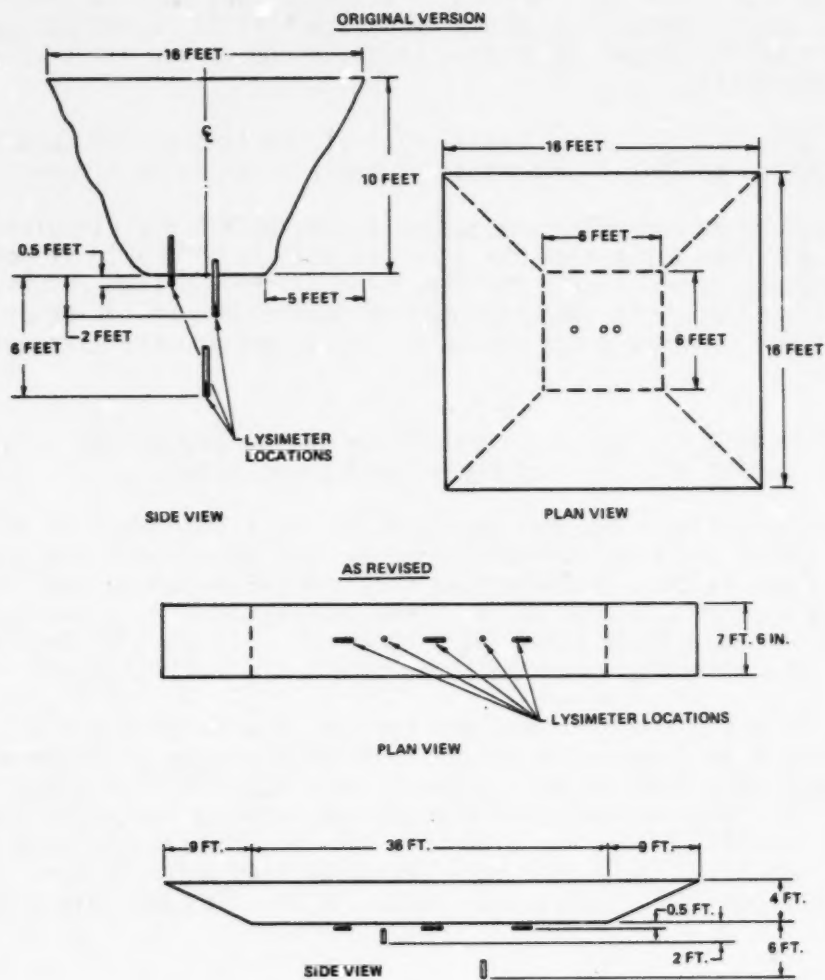


Figure 4 Sludge disposal pits

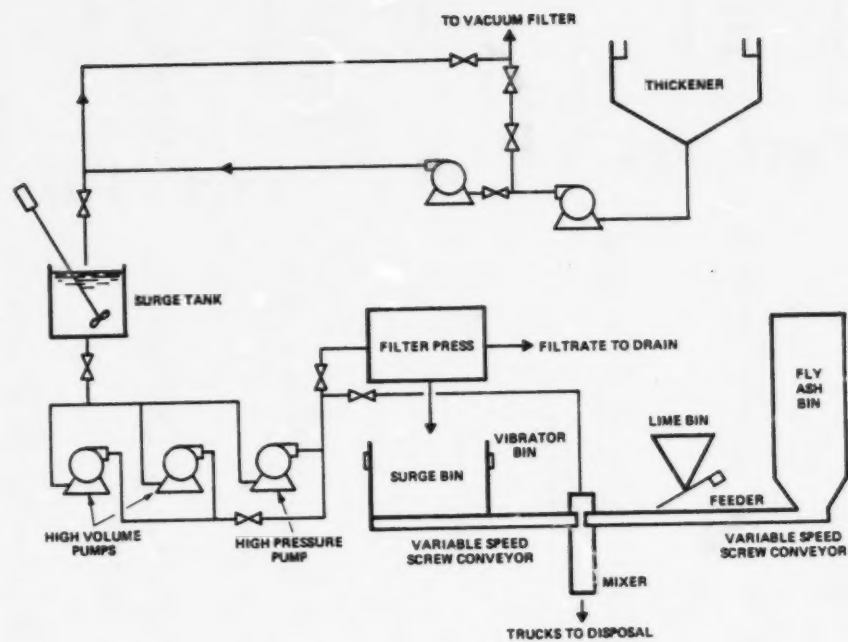


Figure 5 Equipment schematic

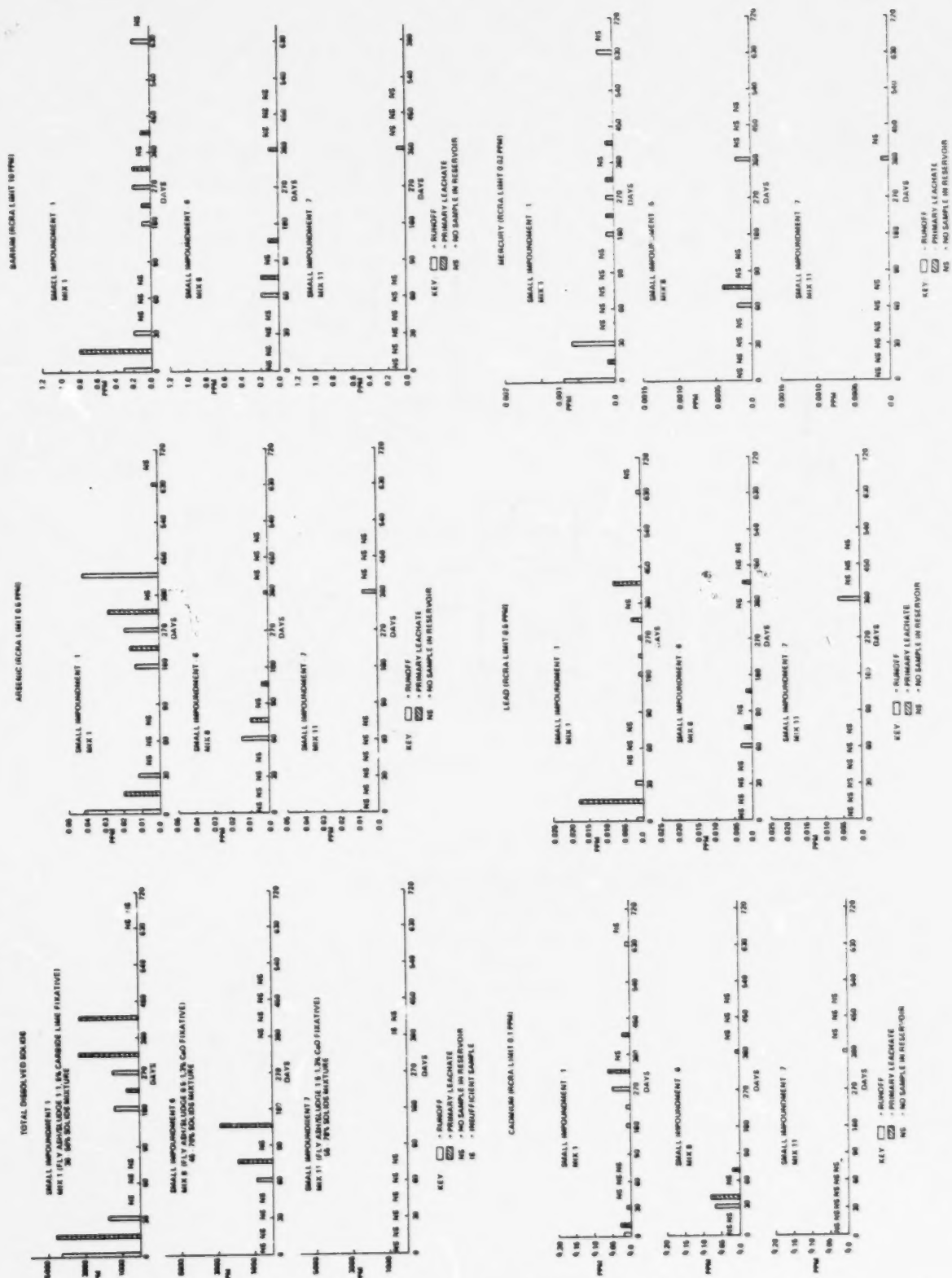


Figure 6 Leachate and runoff samples from small impoundments

These samples were collected from sites with different degrees of stabilization. Mix 1 (small impoundment No. 1 and large impoundment No. 1) was placed as a wet slurry, with a solids content of approximately 42%. No compaction was possible and after several days, the mix had settled leaving a foot of free standing water in the impoundments. The materials remained soft, with little load bearing capacity, throughout the test program.

Mix 8 (small impoundment No. 6) had an initial solids concentration of approximately 57% and represented a higher degree of dewatering of the sludge than Mix 1. This material developed a bearing capacity of >1.5 TSF, as measured by the University of Louisville.

Mix 11 (small impoundment No. 7 and large impoundment No. 5) contained twice as much fly ash as Mix 8 and was placed at an initial solids concentration of approximately 68%. This mixture had a bearing capacity of >>1.5 TSF.

In all cases, Mixture 1 in impoundment 1, a non-compacted mixture formulated using non-dewatered sludge, provided the highest concentrations of trace elements in the leachate. It should be noted, however, that all analytical results were well below the recommended RCRA (Resource Conservation and Recovery Act) limits for trace elements (Federal Register December 18, 1978, section 250.13d).

Note that the leachate and runoff samples from the uncompacted mixture tend to contain high initial concentrations of contaminants which decrease gradually with time (the lack of sample at 60 and 90 days resulted from early freezing as noted earlier). Corresponding samples in the compacted small impoundments 6 and 7 (mixtures 8 and 11) show a much more gradual buildup of trace elements, a lower maximum and a tapering off to a minimum level or no sample after 1-1/2 years.

Sampling of the large impoundment was not affected as severely by weather as in the small above-ground impoundments. The lysimeters were located well below the frost line.

The analysis of a series of leachate samples collected from large impoundments No. 1, 3 and 5 are shown in Figures 7-10 for several major and trace elements. Several points should be noted:

- 1) The quality of leachate improved and the quantity decreased with increasing depth beneath the disposal sites. This is an indication that the filtering action of the soil may aid in decreasing the concentration of contaminants reaching the ground water supply. In fact, no samples were available for collection from the 2-foot or 6-foot sample depths below large impoundment No. 5. This mixture had a permeability of less than 4×10^{-6} cm/s within 60 days after placement.
- 2) The leachate quantity decreased and the quality improved with increased initial dryness of the mix. Mix No. 1, which was placed as a "soup-like" slurry, produced the highest concentration of contaminants measured in the leachates during the program.
- 3) The concentration of contaminants generally decreased with time from those impoundments where a sufficient number of samples were collected to assess the long term trend.

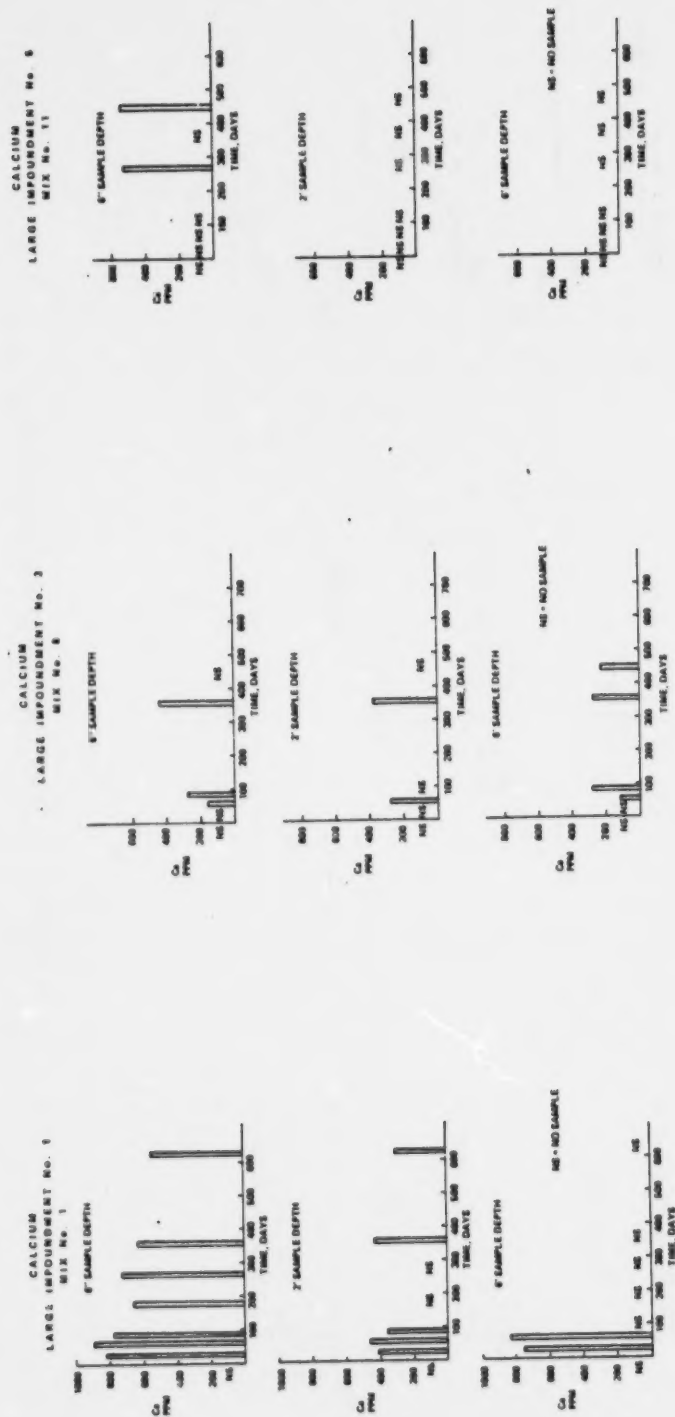


Figure 8 Calcium--large impoundments



Figure 9 Selenium--large impoundments



Figure 10 Chloride--large impoundments

- 4) With only three exceptions, the concentrations of trace elements in all of the leachate samples collected during the program were below the levels established for defining leachates from hazardous wastes under RCRA 250.13(d). The only leachates exceeding these limits were from Mix No. 1 (Selenium and Cadmium) and Mix No. 9 (Cadmium).

Ground water background levels of many priority pollutants approached RCRA limits prior to the start of this demonstration program. Samples obtained from existing wells on the site during the test program have shown no detectable increase in any species currently being monitored. It should also be noted that rain water samples collected during this test program approached primary drinking water standards for several substances.

SUMMARY

Properly prepared landfill from FGD sludge/fly ash mixtures will not contaminate the surrounding groundwater. Results obtained from analysis of leachates from the series of landfill impoundments in this study show that trace elements on the RCRA list of contaminants were found in concentrations below those established to characterize hazardous or toxic waste.

A trend toward decreasing concentrations, with time, of trace contaminants was observed in both leachate and runoff samples obtained from the stabilized sludge mixtures. The small impoundments have provided higher concentrations since no attenuation by local soil is provided and vegetation that might minimize runoff was not established on these sites.

Most sites developed compressive strengths significantly (up to tenfold) greater than the minimum required for recreational or light structural landfill. Water samples obtained from beneath the large impoundments indicate that the filtering action of the soil aids in decreasing the concentration of contaminants reaching the ground water supply. Certain mixtures have undergone a fixation reaction thus minimizing the release of moisture and/or contaminants to the surrounding soil.

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PHYSICAL PROPERTIES OF FGC WASTE
DEPOSITS AT THE CANE RUN PLANT
OF LOUISVILLE GAS AND ELECTRIC COMPANY

by

C. R. Ullrich and D. J. Hagerty
Civil Engineering Department
University of Louisville
Louisville, Kentucky 40208

and

R. P. Van Ness
Louisville Gas & Electric Company
Louisville, Kentucky 40201

Abstract

Physical tests performed on FGC wastes from the Cane Run plant of LG&E included in situ shear strength determinations and in situ plate loading tests, as well as shear strength and permeability tests performed in the laboratory on samples taken from waste deposits at the plant. In situ testing and sampling were done at nominal intervals of 0 days, 30 days, 90 days, 180 days, and 360 days after field placement of wastes. Various mixtures of lime process FGD sludge, flyash and lime (or cement) were tested. EPA-sponsored testing of field deposits disclosed non-homogeneous and discontinuous conditions created by weather effects (freezing and thawing), interruptions in filling and incomplete mixing.

Sample disturbance of the brittle cured wastes was severe; thus, lab strength values were low, about one-half the value of corresponding in situ test strengths. Lab permeability tests on field samples yielded values as much as two orders of magnitude higher than values obtained on lab-cured undisturbed samples. Permeability of field samples varied between 3×10^{-5} cm/sec and 3×10^{-6} cm/sec. In situ shear strength values 30 days after placement varied from about 250 psf to more than 3,000 psf. Strength increases were noted with age for all but one deposit (which was disrupted by freezing). In situ strengths were much lower than strengths of samples cured and tested in the laboratory. Plate loading tests on selected deposits of wastes showed stiff fixed materials with bearing capacities greater than 15,000 psf.

Freezing created layers in waste deposits and retarded or impaired stabilization reactions between lime and wastes. Some dewatering of exposed surfaces (to depths of six inches or so) was caused by freezing and thawing.

Commercial lime was more effective than carbide lime (a waste product available near the Cane Run plant) in stabilizing wastes and performed as well as Portland cement.

PHYSICAL PROPERTIES OF FGC WASTE DEPOSITS AT THE
CANE RUN PLANT OF LOUISVILLE GAS AND ELECTRIC COMPANY

OVERVIEW

To characterize various mixtures of FGC wastes and additives and to investigate their behavior during land disposal, a series of impoundments was constructed at Louisville Gas & Electric Company's Cane Run Plant. After field placement, in situ strength tests were conducted, and samples were taken and were subjected to laboratory permeability tests and unconfined compression tests. Additionally, plate load tests were performed on materials in two test pits.

These physical tests indicated that behavior of mixes of FGC wastes were highly dependent on mix composition. Total solids content is important; increasing solids content from 50% to 67% may increase strength 10-fold while giving a 10-fold decrease in permeability. However, strength gain is highly dependent on the fixative material added to the mix: in these tests, Portland cement was far more effective than commercial lime. Variation of flyash-to-sludge ratio changed the mix behavior but not as drastically as did changes in total solids content and type of additive. High flyash content in a mix changed the material from a somewhat plastic material to a more pervious, friable and brittle material (for these mixes with calcium sulfite sludge from a lime scrubber).

Other factors of great influence were quality of mixing, continuity of placement, weather conditions at placement and weathering effects (freeze-thaw especially) after placement. Poor mixing caused heterogeneity in waste deposits. Interruptions in placement caused the formation of distinct layers (wetter, softer zones between denser, harder layers). Freezing accentuated layering, with consequent changes in solids content, strength and permeability. Hot weather led to rapid reactions and early strength gain; in some cases, some of this strength disappeared during a complete cycle of weathering.

Laboratory tests prior to field placement, conducted on very well-mixed, carefully cured samples, did not forecast accurately the behavior of not so well-mixed materials placed in layers in all extremes of weather. Moreover, extreme care was required to obtain samples from field deposits without seriously disturbing the brittle but very stiff wastes mixes.

These tests showed that mixes of FGC wastes with 3-5% lime or cement may exhibit strengths as high as 5,000 psf (undrained shear) with permeability as low as 10^{-6} cm/sec, if total solids content is high and placement conditions are favorable. Details of these findings are given in the following pages.

INTRODUCTION

This paper presents results of field and laboratory testing done on various mixtures of scrubber sludge, flyash, and lime. The sludge mixtures were contained in impoundments at Louisville Gas and Electric Company's Cane Run Plant, and testing was done by the University of Louisville from November 1976 to October 1978, under contract to LG&E.

Sludge mixtures impounded at the Cane Run disposal site were contained in two basic styles of impoundments: 25-cuyd capacity backyard swimming pools and 50-cuyd excavated pits. Ten swimming pools were erected: four of these were used for sludge mixtures utilizing carbide lime and six for mixtures utilizing commercial lime. Excavated pits were of two styles: 10-foot x 10-foot x 8-foot deep pits and 20-foot x 10-foot x 4-foot-deep pits. Two deep pits were used for sludge mixtures utilizing carbide lime and the other three pits were used for commercial lime-sludge mixtures. Details of the impoundment layout, mix designs, and placement procedures are given in a companion paper by Mohn, et al.¹

DESCRIPTION OF WORK PERFORMED

As requested by Louisville Gas & Electric Company, work done by the University consisted of three basic tasks:

- 1) The local soil/hydrology at the disposal site was studied for purposes of locating tests impoundments.
- 2) In-situ vane shear strength tests were to be performed at nominal intervals of 0 days, 30 days, 90 days, 180 days, and 365 days after placement.
- 3) Laboratory unconfined compression tests and permeability tests were to be performed on undisturbed samples taken at the same time that the in situ vane shear strength tests were performed.

Sampling and testing was done as closely as possible to scheduled times, although inclement weather and other circumstances sometimes interfered with strict adherence to the schedule. Laboratory testing was done usually within one week of sample procurement if at all possible. Because of the expiration of the first phase subcontract, it was not possible to do field and laboratory testing on sludge mixtures containing carbide lime 365 days after placement. Also, testing at 90 days, 180 days, and 365 days after placement was not done on sludge mixtures containing commercial lime.

To satisfy the deficiencies in the initial work program in 1978 University of Louisville investigators conducted a program of additional physical testing.

The additional physical tests included field vane shear strength measurements and sampling of sludge mixtures contained in Pits 1, 2, 3, 4, and 5 and Pools 2, 5, 9, and 10. Laboratory unconfined compression tests and permeability tests were performed on obtained samples whenever possible. Pools 1, 3, 6, 7 and 8 contained mixes identical to those in Pits 1, 2, 3, 5 and 4, respectively, and were not re-tested. The mix in Pool 4 had emerged from the mixing truck as spheres at placement, had soon frozen and disintegrated upon thawing; consequently, tests on Pool 4 were not relevant.

Vane shear strength tests were made at one-half foot intervals in all impoundments where it was possible to insert the field vane shear strength measuring device. Continuous sampling was done on the sludge mixtures contained in the selected impoundments. Samples were obtained by use of 30-inch long by 2-inch diameter Shelby tubes. A drilling rig was utilized in obtaining samples in the pools and pits where the stiffness of the mixture precluded pushing the sampling tubes into the material manually. From each 2-foot length of sample, two unconfined compression test samples and one laboratory permeability sample were prepared. The laboratory unconfined compression test samples were prepared from 8-inch lengths of sample from the ends of each Shelby tube sample. The unconfined compression tests were performed in accordance with ASTM Standard Method of Test D2166. Laboratory permeability tests were performed on five- to six-inch lengths of sample cut from the middle eight inches of each Shelby tube sample. Testing was done on the sawed length of Shelby tube to avoid sample disturbance.

TEST RESULTS FOR EACH IMPOUNDMENT

Pit No. 1

Mix 1 was placed in Pit No. 1 and consisted nominally of sludge (thickener underflow at 24% solids content), with an equal dry weight of flyash, and 5% by dry weight of carbide lime added, to give a nominal final solids content of 39.3%. The pit was filled in truckloads with overnight interruptions in the filling operation, on November 8-11, 1976. After each sequence of filling, the solids mixture settled and supernatant water appeared on top of the deposit. After final filling, the depth of supernatant water reached approximately 10-12 inches before evaporation removed all standing water. After this water evaporated during November and early December, the exposed top of the deposit froze in late December. The frozen crust of the deposit reached thicknesses of at least 12-18 inches during the winter of 1977 and possibly greater thickness during the severe winter of 1978.

Until the most recent sampling effort in September 1978, only one tube sample had been obtained from Pit No. 1, on February 8, 1977 when the sludge had frozen to a depth of several feet. Prior to that time, the extremely soft consistency of the mixture in the pit precluded sampling. On September 28, 1978 the investigators obtained continuous samples from Pit No. 1 for the first time, and a virtually complete record of field vane shear strength with depth also was obtained. During summer 1977, a desiccated crust was found in the upper few inches of material in the pit, above material of varying consistency. When the pit was sampled in September 1978 a crust again was found in the top 12 to 16 inches of the deposit. A stiff layer was found at 22 inches below the top of the deposit, but this stiff layer was underlain by softer materials.

The results of the 1978 vane shear tests in Pit No. 1 are shown in Figure 1, together with results of field vane shear tests performed during 1976 and 1977. Although the nominal solids content of the material in Pit No. 1 was approximately 39% at the beginning of the testing period, evaporation of supernatant water increased the solids content to between 45 and 55%. At the time samples were obtained in September 1978, the average solids content of the materials at depth in the pit was about 58%; in the top 8-12 inches of material, a dried-out crust had formed with a solids content of approximately 75%.

The tube sample which had been obtained from Pit No. 1 in February 1977 melted when it was returned to the laboratory. The samples secured during the latest sampling operation showed unconfined compression strengths between 250 and 830 psf, with corresponding undrained shear strengths between 125 and 415 psf. The sample obtained in the dry desiccated crust crumbled when it was removed from the sampling tube. The strain at failure in these unconfined compression tests varied from 6.5% to 8% indicating that the material behaved in a more plastic fashion than did much of the material in other pits. Laboratory tests on this mix had shown 60-day unconfined compression strengths inadequate for sample integrity (too soft to test).¹

Permeability tests were performed on two samples obtained from Pit No. 1. The sample obtained from a depth of 6-12 inches in the deposit exhibited a permeability coefficient of 2.3×10^{-5} cm/sec at a solids content of approximately 62%, while the sample obtained between 27 inches and 33 inches below the surface of the deposit showed a coefficient of permeability of 2.9×10^{-5} cm/sec at a solids content of approximately 56%. Permeability tests on this mix conducted prior to field placement had shown a coefficient of permeability of 7.6×10^{-5} cm/sec.¹

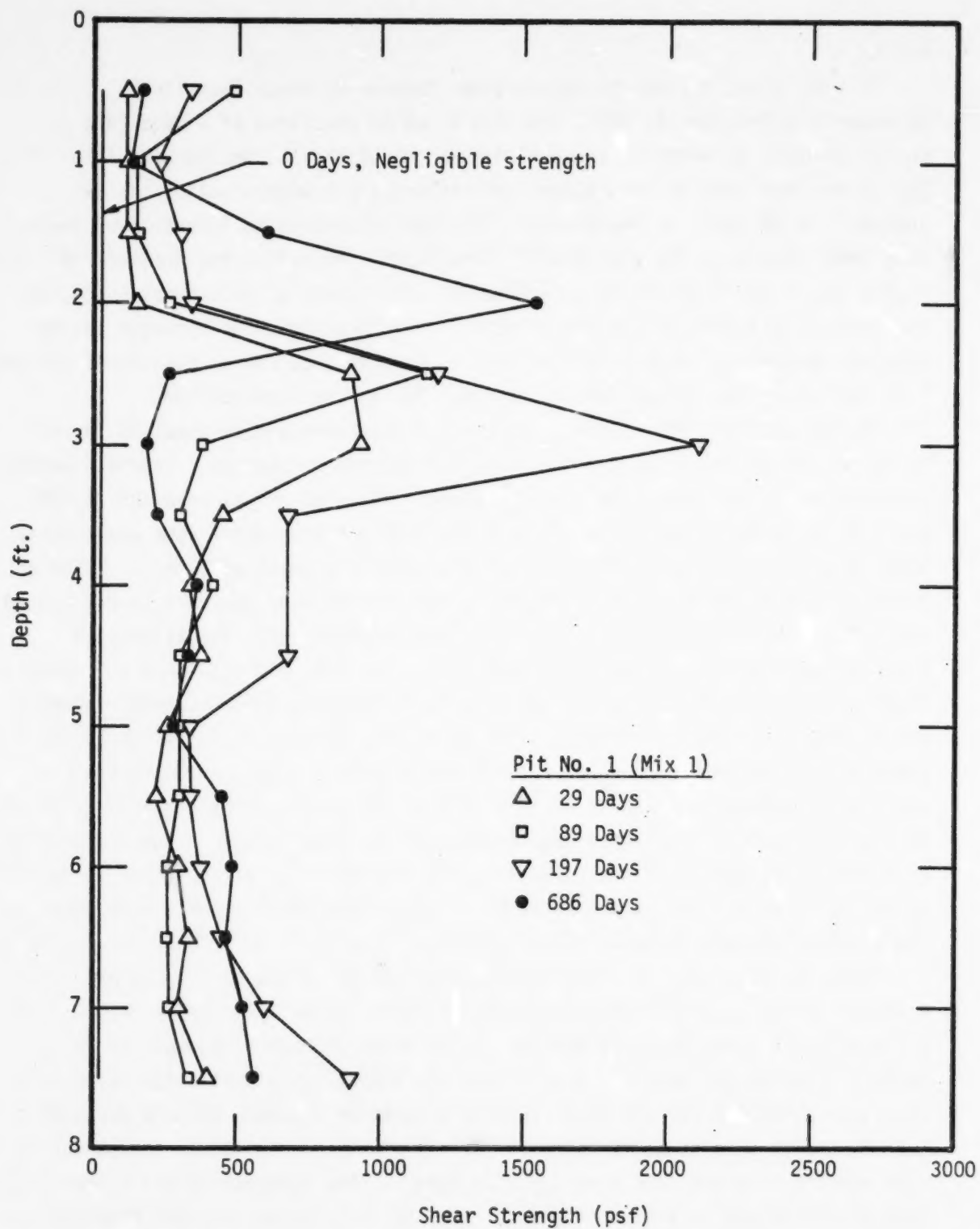


Figure 1 In-Situ Shear Strength Values for Pit No. 1

Pit No. 2

Pit No. 2 was filled, in intermittent fashion by truckloads, from December 2 to December 6, 1976, with Mix 2, which consisted of sludge (42% solids content, by weight), an equal dry weight of flyash, and carbide lime (5% of the dryweight of the sludge) and yielded a mix with a solids content nominally of 59.75%. On December 8, 1976, the resistance to insertion of the vane shear device varied considerably with depth, indicating the presence of layers with a stiff consistency interbedded with layers of soft consistency, in the upper 3 to 4 feet of the pit deposit. The undrained shear strength values obtained during the first investigation in December 1976 varied from about 300 psf near the top of the sludge deposit to about 700 psf near the bottom.

By mid-January, the sludge mixture had frozen downward to about 30 inches below the top of the deposit. The values of undrained vane shear strength varied erratically in the upper 4 feet of the deposit from 750 psf to more than 3,000 psf. A very high value was obtained in the wholly frozen upper zone and another high value (greater than 3,000 psf) was obtained at a depth of 3 feet. Below a depth of 4 feet, however, the strength varied much less dramatically and averaged about 7000 psf. When in situ tests were done in March 1977, the effects of freezing again were apparent to a depth of about 4 feet although some surficial thawing had occurred. Below a depth of 4 feet, the vane shear strength values varied from about 500 psf to about 1,000 psf. Again, a stiff layer was indicated at a depth of about 3 feet. In late June 1977, in situ strength tests again were performed. Below a depth of 4 feet the shear strength varied from 600 psf to 1,200 psf, slightly greater than the in situ values measured 100 days after placement. In the upper 4 feet of the deposit, strength values in excess of 3,000 psf were measured at depths of 2 feet and 3.5 feet, with much lower values measured between these depths.

When the site was revisited on September 28, 1978, the in situ shear strength in the upper 4 feet of the deposit again varied erratically from a value of about 1,900 psf near the surface to about 800 psf at a depth of 18 inches. Hard layers were found at 2 feet and 3 feet, with a strength value in excess of 3,000 psf for the hard layer at a depth of 3 feet. Below a depth of 4 feet, the in situ strength values varied in a pattern similar to the variation noted during previous sampling. In general, the strength values ranged from about 550 psf to a high of about 1,200 psf, in a rather regular fashion. The strength values between 4 feet and 5 feet depth were somewhat lower than the strength values measured in 1977, but below 5 feet the strength values

measured in 1978 generally were somewhat higher than the strength values measured previously. The vane shear test results for Pit No. 2 are shown in Figure 2. In comparison, laboratory tests on Mix 2, conducted prior to field placement, had shown an undrained shear strength of about 4,100 psf.¹

When the investigators visited Pit No. 2 on December 8, 1976, the sludge mixture was too soft to be retained in the tube sampler. In mid-January 1977, samples obtained were completely frozen and required thawing in the laboratory before testing. The strength values obtained in unconfined compression tests on the thawed samples varied from about 250 psf to about 500 psf, and the moisture contents of the thawed samples were considerably lower (38-64%) than the moisture content nominally set for the mixture (67%). On March 15, 1977, when samples were obtained, 99 days after placement, the sludge mixture appeared to be wholly or partially frozen to a depth of about 4 feet. Samples obtained at this time were allowed to thaw slowly in the lab and were tested in June, 177 days after placement. Strength values from these tests varied from 400 psf to 750 psf and moisture contents for these samples varied from about 36% to 49%.

Samples again obtained from Pit No. 2 on June 28, 1977, showed undrained shear strengths between 500 psf and 2,100 psf, with a high value found for a sample taken from a depth of 2 feet. This testing date, July 2, was 208 days after placement. Moisture content values varied from 40% to 59%.

On September 28, 1978, samples again were obtained. When these samples were tested in the laboratory 672 days after placement, the undrained shear strength values varied from 184 psf to 1085 psf. Solids contents ranged from a high of 79% to a low of 65%. These last shear strength test results indicated some deterioration of the near-surface layers of the mixture but at depth the strength appears to have increased slightly with time compared to initial values obtained. The very hard layer found at a depth of 2½ feet and represented by a very strong sample tested 208 days after placement, was not represented in the laboratory shear strength values found in 1978. The values of laboratory shear strength determined for samples from Pit No. 2 are shown in Figure 3.

Whenever samples were obtained from Pit No. 2, laboratory permeability tests were performed on the obtained samples. The first group of samples obtained from Pit No. 2 were tested from 87-113 days after placement of the material. In general, these samples showed permeability coefficients between 4×10^{-6} cm/sec and 2×10^{-5} cm/sec. Samples tested between 193 and 195 days after placement showed permeability values between 4.5×10^{-6} and 6.5×10^{-6} cm/sec except for one disturbed sample taken from a very shallow depth which

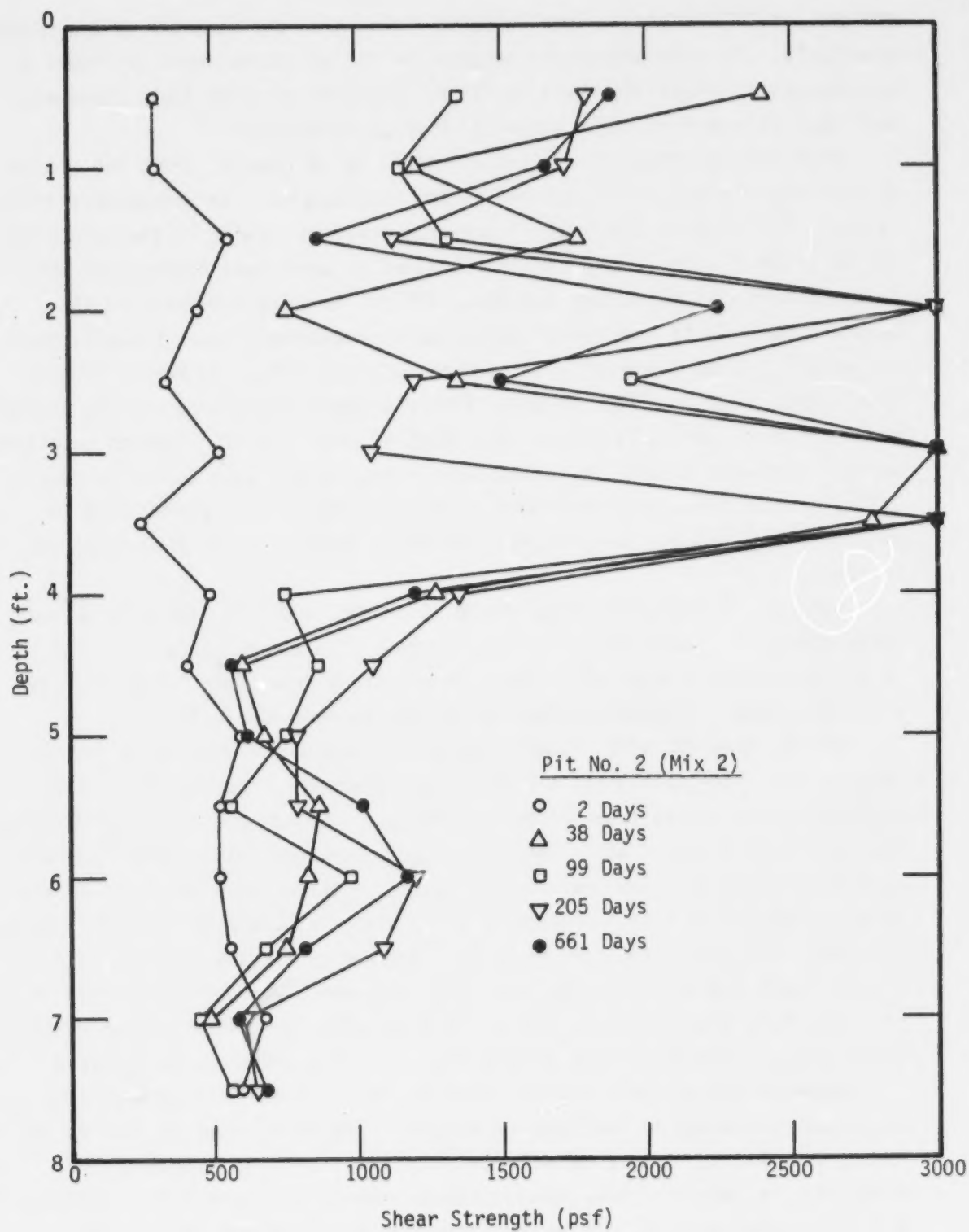


Figure 2 In-Situ Shear Strength Values for Pit No. 2

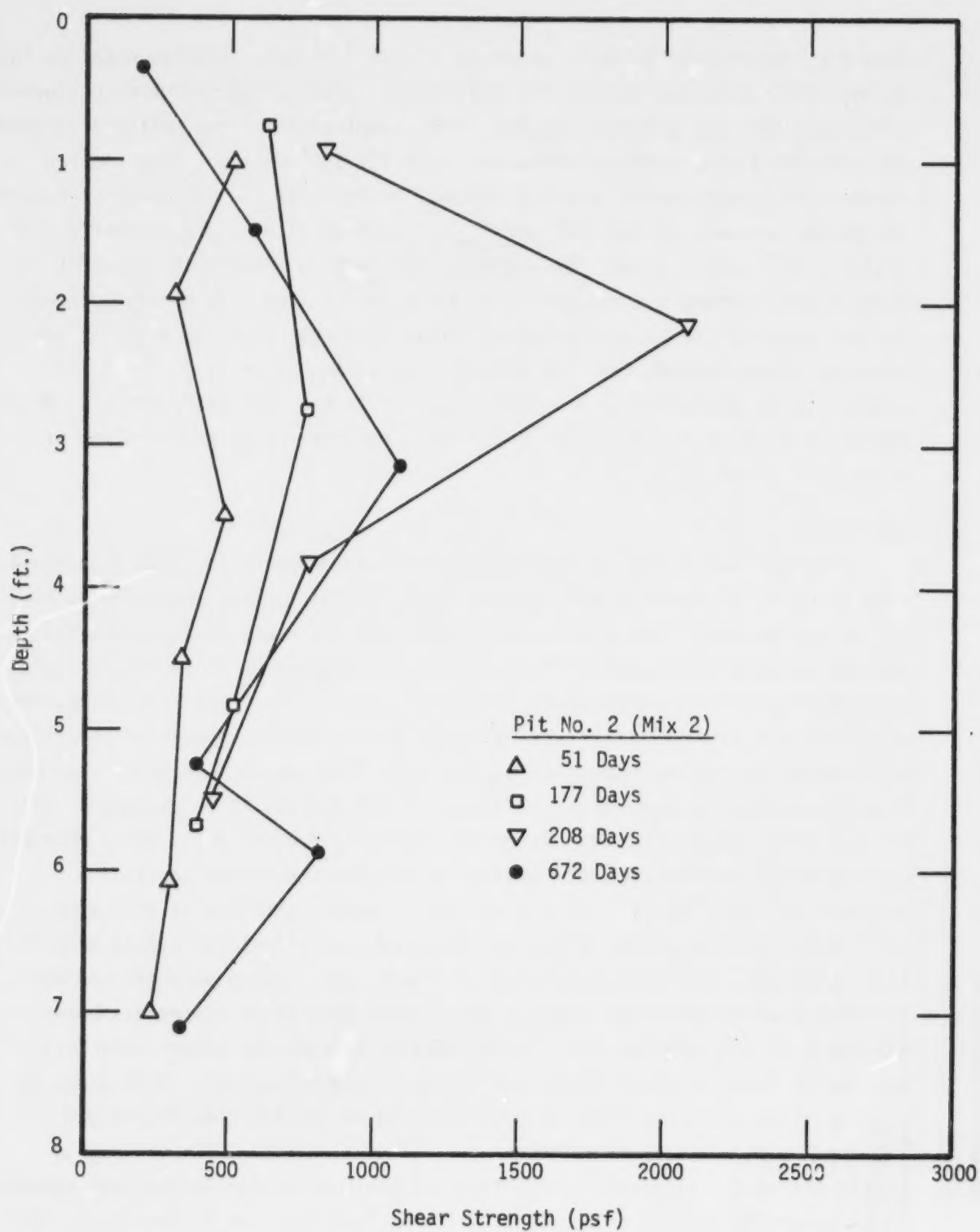


Figure 3 Laboratory Shear Strength Values for Pit No. 2

showed a permeability value in excess of 2×10^{-5} cm/sec. Samples obtained later during 1977 and tested between 270 and 284 days after placement were considerably more disturbed than previous samples; these samples showed low moisture contents and relatively high permeability values of $2-3 \times 10^{-5}$ cm/sec. Four samples taken in fall 1978 showed moisture contents between 39% and 58%, with corresponding solids contents between 63% and 72% and permeability values generally less than 6×10^{-6} cm/sec except for a sample taken within the top 12 inches of the deposit which showed a permeability of 4.5×10^{-5} cm/sec. In general, except for the upper 18 inches of the deposit where disturbance of the material has occurred through weathering, the permeability of the material in Pit No. 2 appears to be approximately 6×10^{-6} cm/sec or less. Laboratory tests on Mix 2 performed prior to field studies had shown a coefficient of permeability of 2.9×10^{-6} cm/sec.¹

Pool No. 2

This pool was filled between November 22 and November 24, 1976 with Mix 4, consisting of thickened sludge (nominally 55% solids content) and carbide lime (5% by dry weight). The final solids content of Mix 4 was 56.2%, nominally. Several weeks after placement the mixture was compacted by blows and pressure from the bucket of a rubber-tired tractor excavator. Prior to compaction, the mixture had a very irregular surface in the pool. Before compaction the sludge was sampled and tested in situ on September 1, 1976, seven days after placement. Samples obtained on that date were tested in the laboratory on January 5, 1977, 42 days after placement. The sludge had partially frozen during early December. The compacted, partially frozen sludge was sampled and tested again on December 24, 1976, 30 days after placement. Samples obtained at this time were tested on January 14, 1977 after they had been allowed to slowly thaw in the laboratory. At seven days after placement, the in situ strength was 500 to 700 psf while the thawed samples taken seven days after placement showed strengths of only 200-250 psf. The difference between the values apparently was due to freezing which gave a false high strength reading. Thirty-day in situ tests on partially frozen sludge gave higher results than the values obtained on thawed samples tested 51 days after placement (750 psf in situ vs 250 psf on lab samples). The entire thickness of sludge in the pool appeared to be frozen by the end of February 1977. At that time the frozen sludge had an apparent strength of from 1,200 psf to 2,100 psf, while the badly disturbed samples had negligible strength after they were thawed in the laboratory 125

days after placement. Testing on June 28, 1977 showed an in situ strength of 800 psf to 1,200 psf with the higher value recorded near the surface of the deposit. A sample tested in the laboratory on July 2, 1977 showed a strength slightly less than 500 psf. This material was quite sensitive to sampling disturbance. Laboratory tests on Mix 4 prior to field placement had shown shear strength of 4,600 psf.¹ When the pool was sampled on September 19, 1978, the in situ strength values varied from 223 psf to 186 psf with the higher value recorded closer to the surface of the deposit. This is a drastic reduction in strength 664 days after placement compared to the strength values obtained at earlier times and compared to the initial pre-placement laboratory test results. The material had been frozen at the time tests were performed 90 days after placement and the high values obtained at that time are not relevant. However, when the pool was tested 216 days after placement, the material had completely thawed and the strength values obtained at that time served as a basis for comparison for the values obtained in September 1978. It appears that a drastic reduction in strength of the material has occurred as a result of weathering during the intervening year's time.

Samples obtained from the pool on September 19, 1978 were tested on September 24, 1978 in the laboratory in unconfined compression. These samples showed shear strengths slightly higher than the in situ strength values obtained at the site: from 196 psf to 336 psf. This is the only instance in the testing program in which the laboratory shear strength values were higher than the in situ vane shear test values. In all other cases, the laboratory values were lower than the in situ values, primarily because of disturbance of the material during the sampling operation, during the transport of samples to the laboratory and during the preparation of the samples in the testing devices. These test results appear to indicate that the material has deteriorated in the field to the point where weathering effects have severely disturbed the mixture and greatly reduced the strength of the material. The moisture content and corresponding solids content of the materials taken from Pool No. 2 indicate that the materials have not changed appreciably in moisture content since they were placed in November 1976.

Permeability tests were performed on samples obtained from Pool No. 2 as soon as those samples could be thawed and prepared for the tests. The values of coefficient of permeability for test samples taken nominally at 30 days and at 90 days after placement varied from about 0.7 to 1.6×10^{-5} cm/sec. A test was run on a sample obtained on September 19, 1978 and the corresponding permeability value obtained for that sample was 5.2×10^{-6} cm/sec. Laboratory tests

on Mix 4 prior to field placement had indicated a coefficient of permeability of 4.5×10^{-7} cm/sec.¹

Pit No. 3

Pit No. 3 was filled between July 13 and July 16, 1977, intermittently with truckloads of Mix 8 consisting of sludge (50% solids content), an amount of flyash equal in weight to one-half the dry weight to the sludge solids, and commercial lime in an amount equal to 3% of the dry weight of the sludge solids. This mix had a nominal solids content of 60.5%. On July 19, three days after finish of placement, samples were obtained and in situ strength measurements were made. The in situ strengths varied from about 100 psf to more than 3,000 psf. The very stiff zone at a depth of 4.5 to 5 feet corresponded approximately to the surface of the first day of filling of the deposit. Exposure at the top of this deposit to sun and open air during July before the remainder of the mix was placed caused drying with subsequent lime-flyash reactions and hardening. The samples taken on July 19 were tested on July 29 and August 1 in the laboratory. The sample taken from a depth of 12 inches crumbled when it was extruded. Samples from 6 and 6½ feet depth were cracked and disturbed and could not be tested. Tests on the remaining samples showed strengths varying from about 1,300 psf to 1,800 psf at moisture contents of 40% to 53%. The results of these tests, 15 days after placement, are shown in Figure 4. Samples again were obtained on August 24, 39 days after placement, but only with great difficulty. It was not possible to insert the vane shear strength device into the deposit at that time. Samples were obtained only by driving the sample tubes into the mixture with a 16-lb sledge hammer. The samples so obtained were severely disturbed. When these samples were tested on September 2-3, 48 days after placement, shear strength values ranged from 200 psf to 2,200 psf. The high value was obtained on a sample from a depth of about 18 inches with a moisture content of about 37%. The remainder of the samples had apparent strengths of 750 psf or less. These samples contained cracks and voids from the disturbance created while they were being obtained. Laboratory tests on Mix 8, prior to field placement had shown a 60-day undrained shear strength of 3,650 psf.¹

On October 6, 1978 an attempt was made to obtain in situ vane shear strength values from this deposit but it was impossible to insert the vane shear device to a significant depth in the deposit. The surface of the deposit was disturbed, apparently from the effects of weathering, and the dry material was very crumbly. The samples obtained at this time were tested in the laboratory

461 days after placement in the field. The shear strength values for these samples ranged from 127 psf to 3,034 psf. The very low value was obtained on a sample which split and crumbled as it was being tested in the preliminary stages of the test. This sample is not considered to be truly representative of the material in the deposit.

The upper 3 to 4 feet of this deposit appear to have deteriorated somewhat through the effects of weathering since the values of strength of the samples obtained in October 1978 are lower than values of strength obtained at earlier dates at the same depths. However, below a depth of about 4 feet the material appears to be intact and the shear strength in the most recent test varied from slightly more than 1,500 psf to more than 3,000 psf. These samples did not display the disturbance which was apparent in the samples tested earlier because a drilling device was utilized in obtaining the later samples. This device eliminated much of the disturbance associated with hammering a sampling tube into the sludge deposit. Wherever possible, sampling tubes were rotated into the sludge deposit, or they were pushed with a steady smooth push into the material rather than being hammered. The results of the strength tests are shown in Figure 4 where laboratory shear strength values are combined with field shear strength values.

Two samples obtained on July 19, 1977 from Pit No. 3 were subjected to permeability tests on September 3, 1977, 49 days after placement and on September 14, 1977, 60 days after placement. These two tests gave indicated values of coefficient of permeability of 7×10^{-6} cm/sec and 2×10^{-6} cm/sec at moisture contents of 26% and 37%, respectively. Three samples taken on October 6, 1978 also were subjected to permeability tests. These samples had moisture contents between 41 and 45%, or solids contents between 69% and 71%. These solids contents are appreciably higher than the solids content nominally established at the time of placement (60%). The samples were somewhat dry and contained some very small cracks. These samples yielded values of coefficient of permeability of 1.4×10^{-5} cm/sec, 5.9×10^{-5} cm/sec, and 0.9×10^{-6} cm/sec. The lower value of permeability coefficient obtained on the sample from a depth of 65-70 inches in the deposit may reflect a lower permeability for the material in this pit below the depth of material affected by weathering. Laboratory tests on Mix 8 prior to field placement had shown a coefficient of permeability of 4.1×10^{-6} cm/sec.¹

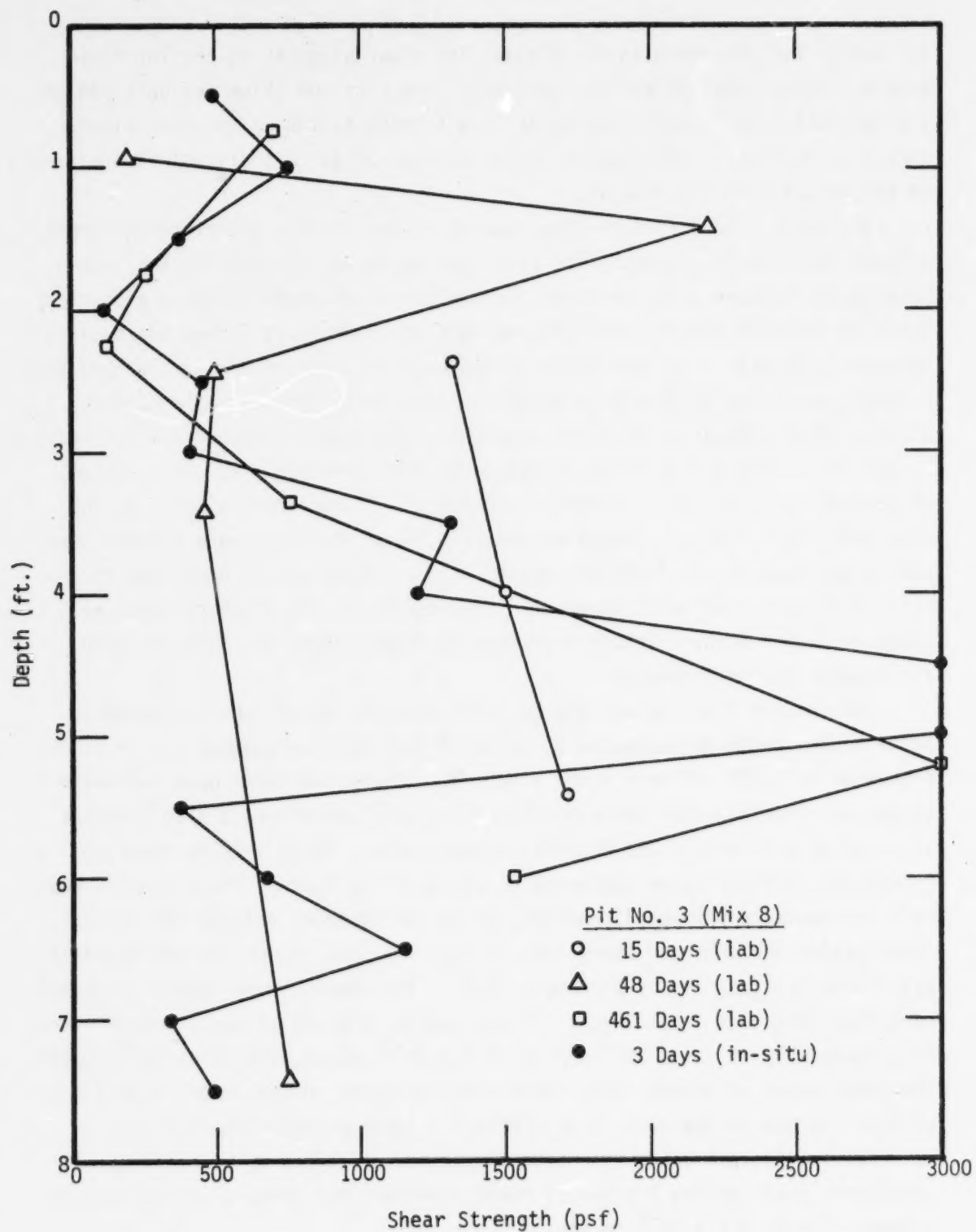


Figure 4 Shear Strength Values for Pit No. 3

Pool No. 5

Pool No. 5 was filled from July 8 to July 11, 1977 with Mix 9, a mixture of sludge (50% solids content), commercial lime (3% by weight of the dry weight of the sludge solids), and flyash (150% of the dry weight of sludge solids). The mixture had a nominal solids content of 71.7%. On July 11, samples were obtained and in situ shear strength measurements were made. The strength values varied from 400 psf to about 650 psf. A sample tested in the laboratory next day showed only slight disturbance and yielded a strength value of about 550 psf. Laboratory tests prior to field placement had shown a 60-day shear strength of 6,250 psf.¹

On August 11, 1977, 31 days after placement, the sludge had hardened sufficiently that the vane shear strength device could not be inserted into the deposit. Samples were obtained by driving thin-walled sampling tubes into the sludge. When these samples were extruded from the tubes in the laboratory on August 22, it was seen that the samples had cracked and split badly. The high flyash content of this mix made the material extremely brittle and susceptible to disturbance. On September 19, 1978, it was impossible to obtain a reading of shear strength in situ because the vane shear device could not be inserted. This high flyash mixture which had been exposed to the elements was disturbed only on the uppermost surface. Several samples were obtained and returned to the laboratory for testing. Prior testing had indicated a shear strength value of 550 psf on a sample taken on the day of placement. Samples taken on September 19, 1978 were tested 445 days after placement; two of these samples were severely disturbed. One sample yielded a strength value of 2,770 psf; the solids content for this sample was 69%, very close to the initial solids content established for the pool.

A sample taken on July 11 was tested on September 9, 1977, 59 days after placement, for coefficient of permeability. The indicated coefficient of permeability was 4.5×10^{-6} cm/sec, compared to a coefficient of 5.7×10^{-7} cm/sec obtained in the laboratory prior to field placement on undisturbed material.¹

The samples taken from Pool No. 5 during September 1978 were so friable and brittle it was not possible to prepare a sample for permeability testing in the laboratory.

Pit No. 4

Pit No. 4, only 4 feet deep compared to the 8-foot deep earlier pits, was filled during the period of July 25 to July 31, 1977 with Mix 12 consisting of sludge (50% solids content), an equal dry weight of flyash, and calcium

hydroxide equal to 3% of the dry weight of the sludge solids. The nominal solids content of the mix was 67%, with a moisture content of 50%. Samples were obtained and in situ shear strength values were measured immediately after the pit was filled. The in situ strength values varied from 700 psf to 2700 psf as is shown in Figure 5. A significant variation in resistance to penetration of the vane shear device at the time of the testing indicated definite layering in the deposit. Samples taken on July 31, 1977 were subjected to unconfined compression tests on August 1 and 2. Some sample disturbance was noted, but it was not severe. Shear strength values obtained from the unconfined compression strength tests ranged between about 500 psf and 1,000 psf, from one-half to one-third of the in situ strength for the same depths. The lower strength values resulted from the inevitable sample disturbance that was created when sampling tubes were hammered into the brittle but stiff mix in Pit No. 4. Samples selected for permeability tests also proved to be unsuitable because of the presence of cracks and voids in the samples.

On August 31, 31 days after placement, it was impossible to insert the vane shear device into the hardened sludge. Samples were obtained only with great difficulty. Laboratory strength tests on September 3 and September 6, 1977 showed shear strength values of from 3,000 psf to 1,300 psf for samples with moisture contents of 36% and 37%, respectively. This moisture content was significantly below the placement moisture content of 50%. The lower strength was obtained on a sample from 30-34 inches while the higher strength was obtained for a sample from a depth of 9 to 13 inches. Cracks throughout the samples precluded permeability testing.

This pit was sampled and tested again on October 6, 1978. At that time, the surface of the deposit was too hard to permit inserting the vane shear test device. Samples were obtained by use of the drilling rig. The sample tubes were filled by pushing the tubes into the deposit or by rotating the tubes into the deposit with the drilling rig. These samples were subjected to unconfined compression strength tests 452 days after placement. The moisture contents for the unconfined compression test samples varied from 41% to 51%. Many of the samples contained crevices and lateral cracks. The shear strength values ranged from a low of 492 psf for a sample taken between 2 inches and 6 inches depth in the deposit, to a high of 2,034 psf for a sample from 23 to 27 inches depth. The corresponding solids content for these samples was 68%. Laboratory tests on samples of Mix 12 prior to field placement had shown a 60-day shear strength of 6,850 psf.¹

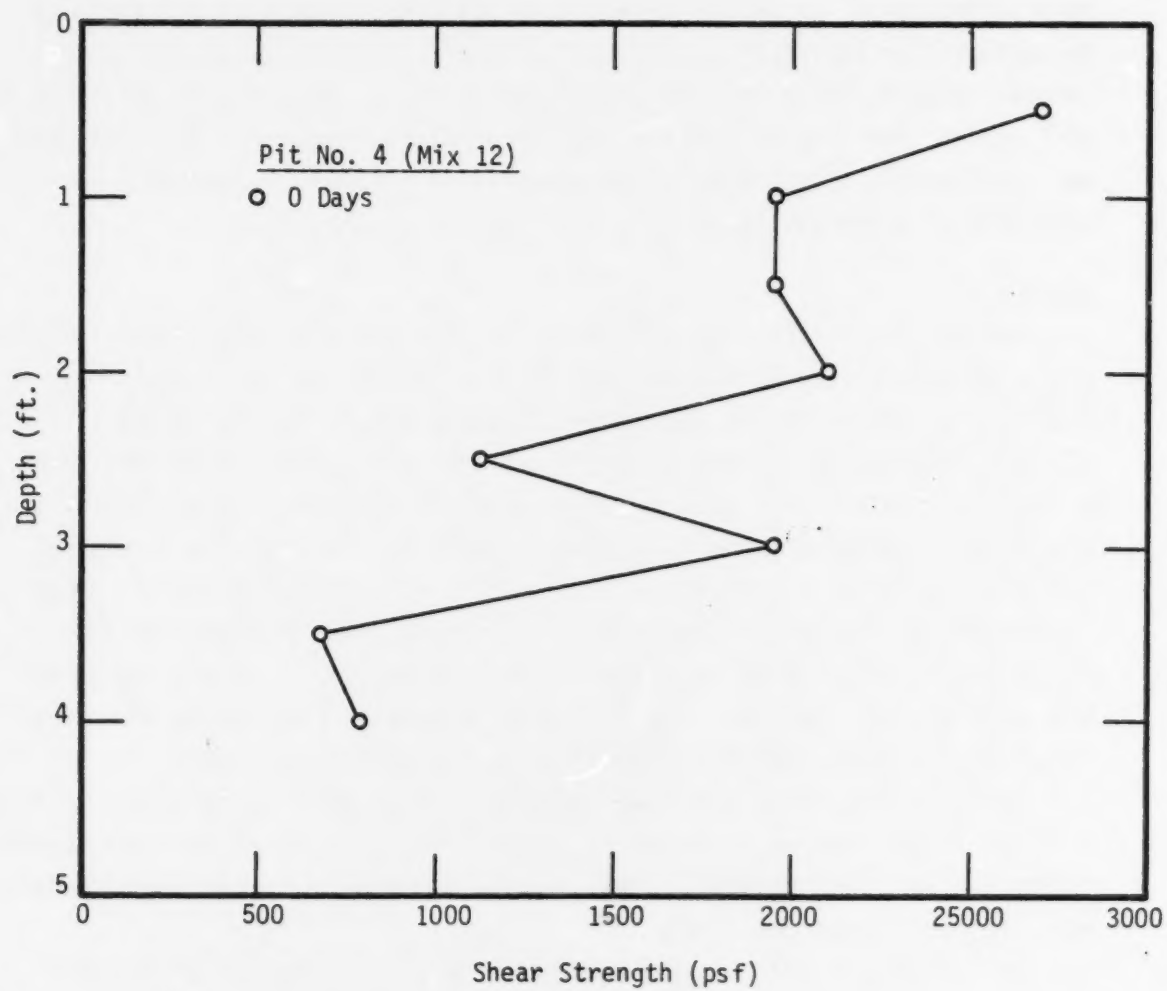


Figure 5 In-Situ Shear Strength Values for Pit No. 4

Because of disturbance during sampling in the program carried out during 1977, it had not been possible to perform permeability tests on samples taken from Pit No. 4. The samples taken from Pit No. 4 on October 6, 1978 did not exhibit the high degree of disturbance associated with the samples taken earlier, although some of the samples had small cracks and crevices. These samples were subjected to laboratory permeability tests 473 days after placement of the material in the field. The resultant coefficients of permeability for samples from Pit No. 4 were 3.0×10^{-6} cm/sec, 3.3×10^{-6} cm/sec, and 1.8×10^{-6} cm/sec, for samples with corresponding solids contents of 68%, 71%, and 68%. Laboratory tests on Mix 12 prior to field placement had shown a coefficient of permeability of 9.2×10^{-6} cm/sec, at 67% solids.¹

Pit No. 5

Pit No. 5 was filled on July 19 to 21, 1977 with the same mixture, Mix 11, which had been placed earlier in Pool No. 7: sludge (50% solids content), an equal dry weight of flyash, and commercial lime (3% of the dry weight of the sludge). The mix was placed in layers and compacted with a shovel device on a tractor excavator. On July 31, 10 days after placement, the surface of the mix could not be penetrated with the vane shear device. Samples were taken but only with great difficulty. These samples were significantly disturbed. Tests on these samples, on August 2, 1977, showed shear strengths of from 750 psf to 2,000 psf with the higher value obtained from a sample taken near the surface of the deposit. The moisture contents of the samples were about 41%, slightly lower than the nominal mix placement moisture content of 50%.

On August 31, 1977, additional samples were obtained. Laboratory tests on these samples showed strengths of 750 psf and 1,750 psf at moisture contents of 39% and 34%, respectively. These samples had been disturbed significantly when they were obtained.

On October 12, 1978 samples were obtained through the use of the drill rig, without a significant degree of disturbance. When these samples were tested in the laboratory 466 days after the material had been placed, one sample showed a very low value of strength of 152 psf, but this sample had been taken between a depth of 1 inch and 5 inches in the deposit and it contained very significant lateral cracks. Another sample taken between depths of 28 and 34 inches split when it was being extruded from the sampling tube. The remaining samples were intact and showed good test results in that no abrupt splitting failure occurred; the test results could be considered representative of the material in the field. The shear strength for these materials varied

from 1,100 psf to more than 1,500 psf at solids contents of approximately 70% (moisture contents between 34 and 43%). Laboratory tests on samples of Mix 11 performed prior to field placement had shown an undrained shear strength of 4,250 psf, at a solids content of 67%.¹

Permeability tests were performed on samples taken on July 31, 1977, 10 days after placement of the material in the pit. These tests were performed 57 days after placement and 90 days after placement. The obtained coefficients of permeability for these samples were 4.5×10^{-6} cm/sec for the 57-day-old sample, and 16.5×10^{-6} cm/sec for the 90-day-old sample. No permeability tests were possible on the samples taken in August 1977 because of severe sample disturbance. The samples obtained on October 12, 1978 from Pit No. 5 were subjected to laboratory permeability tests 484 days after placement of the material in the field. The moisture content of these samples was 45%, with a corresponding solids contents of 69%. These tests showed coefficients of permeability of 1.2×10^{-6} cm/sec for the sample taken from a depth of 12-18 inches and 1.4×10^{-6} cm/sec for the sample taken from a depth of 34-40 inches. Laboratory tests on Mix 11 prior to field studies had shown a coefficient of permeability of 2.9×10^{-6} cm/sec.¹

Pool No. 9

Pool No. 9 was filled on August 1, 1977 with Mix 10, consisting of sludge (50% solids content), an amount of flyash equal in weight to the dry weight of the sludge solids, and Portland cement (3% of the dry weight of the sludge solids). The mix had a nominal solids content of 67% with a corresponding moisture content of 50%.

On August 11, 10 days after placement, samples were obtained and in situ shear strength measurements were made. The strength of the material at all depths exceeded the capacity of the vane shear device so that it was possible only to say the strength was greater than 3,000 psf. Samples were obtained only with great difficulty. One of the samples was tested in the laboratory even though it was cracked slightly. This sample showed a shear strength of 3,250 psf and had a moisture content of 40%, 22 days after placement. All of the samples contained cracks and were disturbed to such a degree that permeability testing was not feasible.

Samples were obtained again on September 21, 1977, 51 days after placement. The material could not be penetrated with the vane shear device. Also, it was very difficult to insert tube samplers into the hardened mix and the samples obtained were disturbed severely. One of the disturbed samples

yielded a shear strength value of 7,445 psf at a moisture content of 40%, when it was tested on October 29, 1977, 79 days after placement. The samples that were tested were cracked but strong; the cracks could not be closed by imposing external pressure. Thus, permeability testing was not appropriate.

When Pool No. 9 was visited on September 19, 1978 again it was not possible to insert the vane shear device into the deposit. However, relatively undisturbed samples were obtained with the drill rig, and these samples were returned to the laboratory and were tested 417 days after placement. These two samples with moisture contents of 37% and 40% showed shear strength values of 5,106 psf and 4,900 psf, at solids contents of 73% and 71%, respectively. Laboratory tests on undisturbed samples of Mix 10 prior to field studies had indicated an undrained shear strength of 2,725 psf, 60-days after sample preparation, with a solids content of 67%.¹

One of the samples obtained on September 19, 1978 from Pool No. 9 was sufficiently undisturbed to permit a permeability test to be performed 427 days after placement. This sample had a moisture content of 43% with a corresponding solids content of 70% and it yielded a permeability value of 3.2×10^{-5} cm/sec. Lab tests on Mix 10 prior to field studies had indicated a coefficient of permeability of 5×10^{-5} cm/sec.¹

Pool No. 10

Pool No. 10 was filled on August 1 and 2, 1977 with Mix 7 consisting of sludge (65% solids content) and flyash in an amount equal in weight to the dry sludge solids; no lime was added to this mix. The nominal solids content of the mix was 78.8%, or 27% moisture content. On August 11, 1977 nine days after placement, samples were obtained and in situ strength measurements were made. The strength values varied from about 650 psf to about 1,300 psf. Samples taken at this time crumbled and broke apart when they were extruded from the sampling tube, on August 23 and 25, 1977. Moisture contents for these samples were 35% and 37%, values higher than the nominal initial mix value of 27%. One of the samples, though cracked and full of voids, supported more than 1,000 psf before it failed abruptly. No permeability tests could be performed on these cracked samples. Sample disturbance was severe; this disturbance was not caused by intense hammering during sampling. Sampling tubes had been inserted without hammering. The high flyash content and lack of cementing agent in this mix made it very crumbly and susceptible to disturbance.

On September 21, the surface of the deposit had hardened to a condition so that the vane shear device could not be inserted. Samples were obtained but only with considerable effort. When these samples were tested on October 3, 1977 and on October 19, 1977, they yielded shear strength values of 1,650 psf and 850 psf, respectively. The samples crumbled somewhat during removal from the sampling tubes and cracked further during test set-up so that no permeability tests were possible.

On September 19, 1978 samples were obtained and it was also possible at that time to insert the vane shear device into the deposit. Field vane shear strengths varied from 2,100 psf to 2,775 psf. Samples obtained at this time were tested in the laboratory, but the very friable and crumbly nature of the material led to significant sample disturbance. One sample with a solids content of 73% yielded a shear strength of 771 psf. The other sample with a solids content of 71% crumbled before it could be tested in unconfined compression. Laboratory tests on Mix 7 prior to field testing had shown a 60-day undrained shear strength of 1,500 psf.¹

Because of the extremely friable and crumbly nature of the samples obtained from Pool No. 10 all during the testing program it was not possible to perform permeability tests on samples from that pool. Laboratory tests prior to field studies had indicated a coefficient of permeability of 7×10^{-6} cm/sec for undisturbed samples of Mix 7.¹

ADDITIONAL TESTING PROGRAM

Because of the high strength results which were obtained from Pit No. 4 and Pit No. 5 during the vane shear testing and during the unconfined compression testing of samples taken from those pits, it was decided that it would be appropriate to conduct plate load tests on those deposits. Consequently, on October 12, 1978, the University of Louisville investigators conducted plate load tests on those two pits. In the plate load tests, a steel loading plate 12 inches by 12 inches was placed on the surface of the mix material in each of the pits after the surface had been swept free of loose debris and had been leveled. Dial gauges were positioned at the corners of the steel plate to measure the downward movement of plate corners under load. An hydraulic jack was utilized to exert downward force on the steel plates used in the loading tests. The results of these loading tests are shown in Figures 6 and 7. In the test on Pit No. 4, a total load of 28,000 lbs was placed on the material to cause a settlement of 1.2 inches. At a settlement of 1.0 inch the pressure exerted by the plate was slightly more than 25,000 lbs/sq ft. This loading

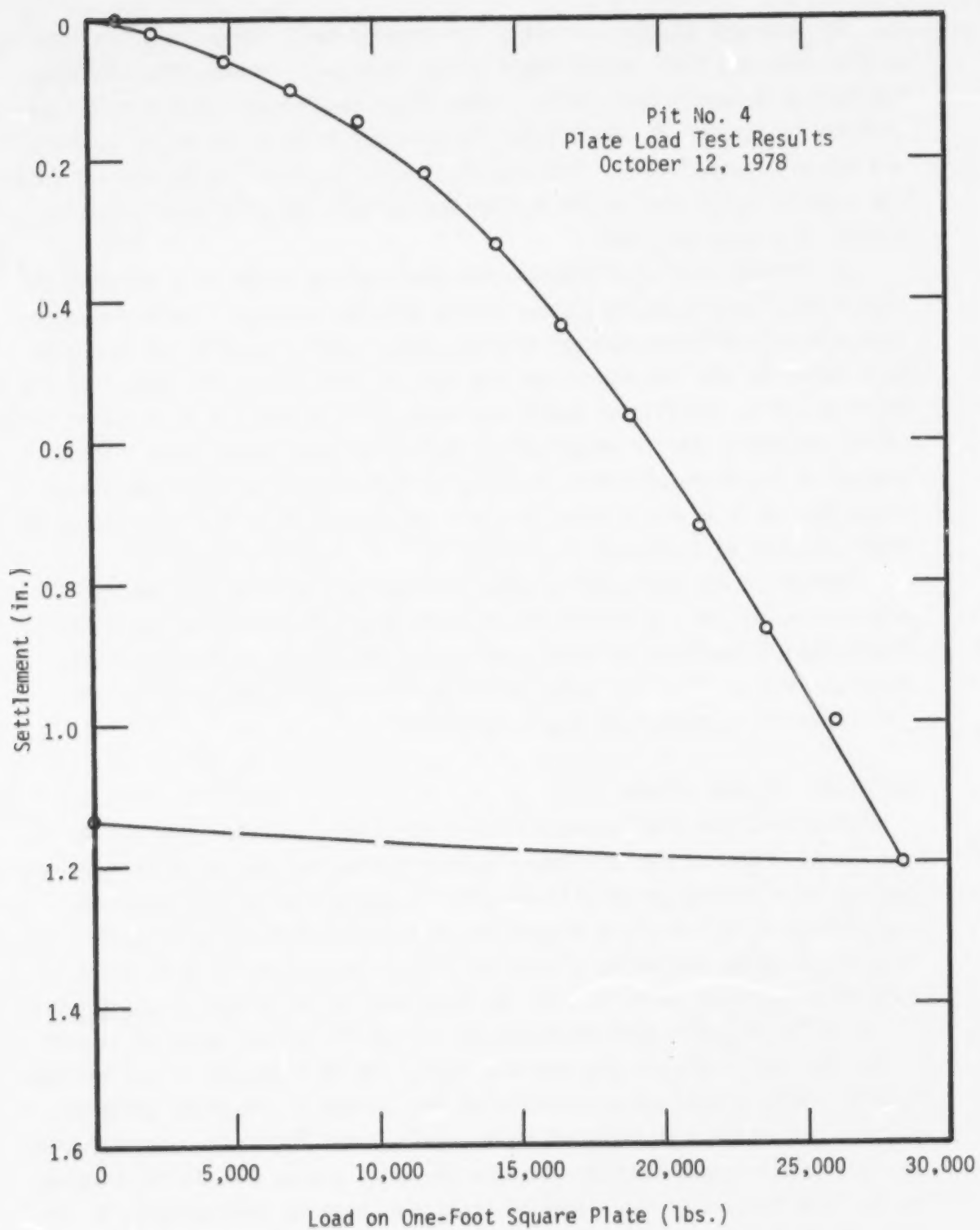


Figure 6 Plate Load Test Results for Pit No. 4

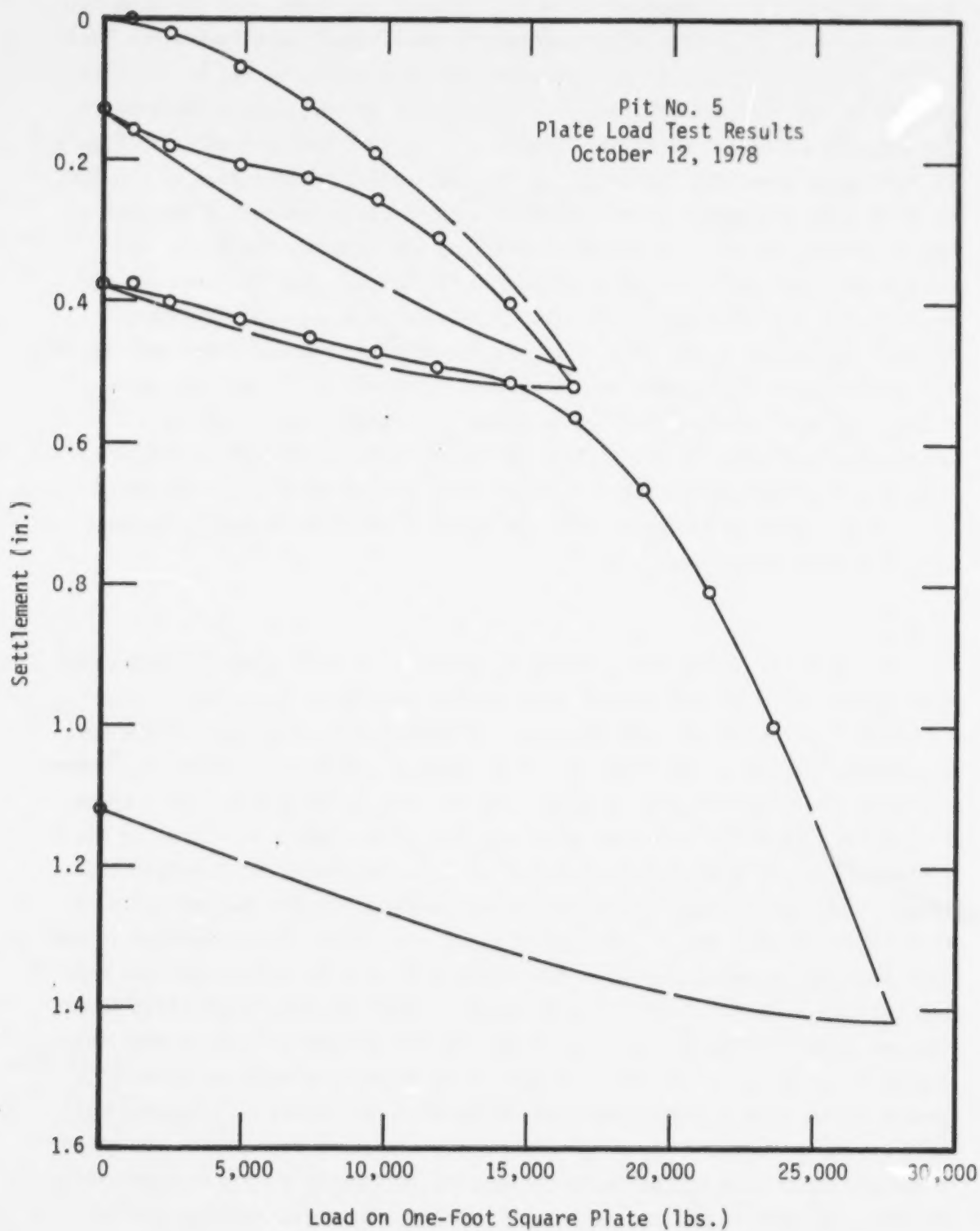


Figure 7 Plate Load Test Results for Pit No. 5

intensity with the accompanying 1-inch settlement indicates that the material in Pit No. 4 would certainly be competent to bear significant foundation loads.

A similar loading test was performed on the material in Pit No. 5, with removal of the load and reloading for two cycles, to investigate the rebound characteristics of the material. The results of this test are shown in Figure 7. It is obvious from the figure that during the loading to more than 16,000 lbs/sq ft with an accompanying settlement of approximately one-half inch, much of the settlement was recoverable; when the load was removed, the plate rose producing a net settlement of only about 0.13 inches. When the load was re-applied, the net settlement under the 16,000-psf pressure again was only slightly in excess of 0.5 inch, but when the load was removed after the second cycle of loading a permanent net settlement of almost 0.4 inches was noted. A third cycle of loading extended to pressures greater than 27,500 psf, and produced a settlement of 1.4 inches. When the load was removed, a net settlement of 1.1 inches had occurred. The behavior of the material in Pit No. 5 under this load test indicates that the material would be competent to bear very high foundation loads.

SYNOPSIS

The material presented in preceding sections of this paper indicates that many factors have influenced the sampling and testing of the sludge-flyash-additive mixtures at the Cane Run plant. These factors also have influenced the interpretation of test results. For example, sample disturbance was severe whenever frozen wastes were sampled; this was particularly true with respect to carbide lime sludge mixtures with very low solids contents. Sampling disturbance also was significant for material that had frozen and subsequently thawed, even in the summer, because the mixture was brittle and sensitive to disturbance in most cases. The use of a drilling rig to secure samples in the 1978 sampling program alleviated this difficulty to a certain extent but did not entirely eliminate sampling disturbance. Additionally, significant layering was noted in pits and pools whenever the filling operations had been interrupted for a period of 12 hours or more; this layering effect was especially severe in the more fluid mixtures and those mixtures placed during extremely cold weather. However, layering effects were also very significant for those materials which were placed in truckloads intermittently during extremely hot weather; in these instances, pozzolanic reactions were accelerated near the surface of the sludge mixtures which were exposed to sunlight and high temperatures, and very stiff layers were produced in some deposits.

Because of the sampling disturbance mentioned above and because of the layering in the deposits, the test data should be viewed with some skepticism. The in situ vane shear strength values are the most reliable data obtained in this investigation. Sampling disturbance reduced the strength of the materials tested in laboratory unconfined compression tests as shown by comparing lab test results for undisturbed samples obtained prior to field studies with those obtained on samples from the field. The effects of sampling disturbance were even more significant in terms of the measured coefficients of permeability obtained in the laboratory. These coefficients of permeability should be considered the maximum values of vertical permeability for the materials. Mass values of vertical permeability in situ may be as much as an order of magnitude lower than the measured laboratory values, because of cracks and fissures in the samples tested in the laboratory. Because of the layering effects mentioned above, mass horizontal permeability is likely to be much higher than mass vertical permeability.

To illustrate the effects of layering and weathering, behavior in Pits 1-5 can be reviewed. Mix compositions for pits/pools are given in Table 1.

With respect to the materials in Pit No. 1, layering, especially in the upper 2 to 3 feet of this deposit, was particularly pronounced. This mix was placed in truckloads with long interruptions between deposits of material. In the upper 2 to 3 feet, the layering may have been caused by poor mixing of the waste materials or by the inclusion of a truckload of much higher solids content material. The effects of freezing may have been accentuated by the presence of layers of more fluid material, created by the intermittent deposition of the wastes in the pit. Ice lenses may have formed when the material froze to a depth of 3 to 4 feet during the winter of 1977. The effects of weathering do not appear to have extended below a depth of about 3 feet. The strength of the material in place today could be taken at no higher than 500 lbs/sq ft for purposes of design. The material is not brittle but is rather plastic and is undergoing very slow consolidation. The strength tests and the permeability tests which were performed on these materials indicate that virtually no cementation has occurred in this deposit.²

In Pit No. 2, the upper 4 feet of the material exhibits severe layering effects. This material froze within the month after it was placed. As in Pit No. 1, it is likely that somewhat wetter areas were formed near the top of each truckload of material that was placed in the pit. This semi-segregation of materials in layers near the top of the pit may have led to the forming of

Table 1 PIT/POOL MIX COMPOSITIONS

<u>Test Site</u>	<u>Sludge Used</u>	<u>Flyash: Sludge</u>	<u>Fixative</u>	<u>Date Placed</u>	<u>Nominal Solids Content</u>
Pit No. 1 Pool No. 1	Carbide lime 24% solids	1:1 (dry)	5% carb. lime	11/11/76	39.3%
Pool No. 2	Carbide lime 55% solids	0:1 (dry)	5% carb. lime	11/24/76	56.2%
Pit No. 2 Pool No. 3	Carbide lime 42% solids	1:1 (dry)	5% carb. lime	12/06/76	59.8%
Pool No. 4	Carbide lime 55% solids	1:1 (dry)	3% carb. lime	12/09/76	71.0%
Pool No. 5	Comm. lime 50% solids	1.5:1 (dry)	3% comm. lime	7/11/77	71.7%
Pit No. 3 Pool No. 6	Comm. lime 50% solids	0.5:1 (dry)	3% comm. lime	7/16/77	60.5%
Pit No. 4 Pool No. 8	Comm. lime 50% solids	1:1 (dry)	3% Ca(OH) ₂	7/31/77	67.0%
Pit No. 5 Pool No. 7	Comm. lime 50% solids	1:1 (dry)	3% comm. lime	7/21/77	67.0%
Pool No. 9	Comm. lime 50% solids	1:1 (dry)	3% Port. cement	8/01/77	67.0%
Pool No. 10	Comm. lime 50% solids	1:1 (dry)	None	8/2/77	78.8%

ice lenses when the materials froze to a depth of 3 to 4 feet below the surface. Ice lensing would have pre-compressed layers of material to yield higher strengths and lower moisture contents. The strength and moisture content values obtained for the upper 4 feet of Pit No. 2 agree with this hypothesis. The materials below a depth of about 4 feet do not appear to have been affected by weathering since the time of placement. The strength of the materials below that depth appear to have increased somewhat in the first two months after placement but they do not appear to have increased significantly since then. The materials in this pit reflect rather poor mixing. The strength of the material would still be insufficient to support sizable structures. The values of coefficients of permeability obtained for the materials in Pit No. 2 varied from about 4×10^{-6} cm/sec to 10×10^{-6} cm/sec, except for values in the upper 18 inches of the deposit where the materials have apparently been severely affected by freeze-and-thaw cycles and other weather effects.

In Pit No. 3, the placement of material verged on random dumping. Large voids were discovered between lumps of the mixture during subsequent testing operations. The material in Pit No. 3 is irregular, contains numerous voids, and could not be considered adequate to support even light structures. Two hard layers were found in the pit between depths of 12 inches and 24 inches and again between depths of 48 inches and 60 inches; these hard layers were apparently caused by the method of placement of the material in the pit. The strength of the material at best could be taken at 1,000 lbs/sq ft for the intact portions of the deposit. Likewise, the intact lumps of material in the pit showed permeabilities of about 10^{-6} cm/sec, but the mass permeability of the deposit would be much higher than that value because of the numerous voids contained in the pit.

The material in Pit No. 4 showed strength values from a high of 3,000 lbs/sq ft near the surface of the deposit to a low of about 1,000 lbs/sq ft near the bottom of the pit, within one day after the material was placed in mid-summer 1977. The material is subject to sampling disturbance as shown by the fact that the unconfined compression strength values of shear strength, obtained on samples tested in the laboratory, amounted to about one-half the value of undrained shear strength obtained with a vane shear device in the field. When the pit was visited in the fall of 1978, a powdery cracked crust was found in the upper few inches in the deposit. Below a depth of 18 inches, this weathering effect was not noted. The undrained shear strength of the material below 18 inches could be taken at a value of 2,000 lbs/sq ft minimum.

The samples taken at 0 days and 31 days after placement were severely disturbed because they were taken by hammering the tubes into the deposit. The samples taken in 1978, approximately 450 days after placement, were taken with a drill rig and were not severely disturbed. The fact that these samples yielded shear strength values somewhat lower than earlier test values indicates that the material in the pit may have weakened somewhat through weathering effects. The values of coefficient of permeability, $2-3 \times 10^{-6}$ cm/sec, were obtained on samples which exhibited very little cracking and disturbance. It is likely that the permeability of the material in the pit is rather close to 10^{-6} cm/sec.

In Pit No. 5, the surface of the deposit has deteriorated with weathering effects evident to a depth of 12 to 18 inches. The samples that were taken 12 and 41 days after placement and later tested in the laboratory were significantly disturbed by the sampling operation, but the samples obtained during fall 1978 were not disturbed in that way. The undrained shear strength of the material in Pit No. 5 could be taken at a minimum value of 1,500 lbs/sq ft, on the basis of the strength tests performed.

Plate load tests indicate that the materials in Pit 4 and in Pit 5 are brittle and tend to gradually yield with a collapsing and compression of voids in the material under load. These materials supported pressures in excess of 25,000 lbs/sq ft during the load tests with measured settlements between 1 and 1.5 inches. The load test results would indicate that a safe bearing value for the materials in Pit No. 4 and in Pit No. 5 would be at least 5,000 lbs/sq ft. These values are considerably in excess of the values obtained through interpretation of the laboratory and field strength tests. The differences between the laboratory and field strength tests, and the results of the plate load tests, reflect the brittle nature of the material. In an undrained shear test, the brittle material tends to be strain-weakening; a failure occurs at a very low value of strain with a loss of strength with subsequent strain. In a compression test such as a plate loading test, the collapse of voids in the material would tend to make the wastes less compressible with increasing strain. In other words, the behavior of the materials in the two tests is different and the results of the two tests are not inconsistent. These materials behave somewhat akin to very weakly cemented aggregates or very weakly cemented natural rock materials, rather than cohesive soils.

Finally, it is important to note the behavior in Pools 9 and 10, since the mix in Pool 9 contains a very effective cementing agent, while the high-solids sludge-flyash mix in Pool 10 contains no fixation additive.

The material in Pool No. 9 is extremely hard at the present time. Very high strength values were obtained in this pool soon after the material was placed. These high strength values indicate the great effectiveness of Portland cement as a binding and stabilizing agent in sludge and flyash mixtures. Samples obtained from Pool No. 9 were disturbed significantly, but even so, the minimum shear strength value for the samples obtained in 1978 was 5,000 lbs/sq ft. A disturbed sample taken late in 1977 gave a value of shear strength in excess of 7,000 lbs/sq ft. The material in Pool No. 9 also showed very little evidence of deterioration from weathering effects; the surface of the deposit consists of a very hard crust. The permeability of the materials in Pool No. 9 should not be greater than the value of 3×10^{-6} cm/sec obtained on a disturbed sample from that pool. It is likely that the in situ permeability of this pool is considerably lower than that value.

The material in Pool No. 10 is extremely friable, exhibiting very little permanent cementation. The shear strength values of the material in Pool No. 10 varied from 700 to 1200 lbs/sq ft nine days after placement, and 30 days after placement the material was too hard to penetrate with the vane shear strength apparatus. Samples obtained in 1978, and tested more than 400 days after placement, showed shear strength values (in laboratory tests) in excess of 2,000 lbs/sq ft. The undrained shear strength of the material could be taken conservatively at a minimum value of 2,000 lbs/sq ft. It appears that the surface of the deposit, in the upper 12 to 18 inches of the material, has deteriorated somewhat through weathering effects. The material is extremely sensitive to disturbance such as occurred when tubes were hammered into the deposit in order to obtain samples. All the samples obtained from Pool No. 10 contained cracks and fissures in such a degree that no permeability tests could be run on samples from that pool.

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SUMMARY OF UTILITY DUAL ALKALI SYSTEMS

by

Norman Kaplan
Industrial Environmental Research Laboratory
Office of Research and Development
Environmental Protection Agency
Research Triangle Park, North Carolina 27711

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ABSTRACT

The Environmental Protection Agency (EPA) has been actively involved in the progressive development of dual alkali flue gas desulfurization (FGD) technology in the U.S. from bench-scale to pilot plant to prototype to full-scale application. This logical progression has allowed a better understanding of the chemistry and made possible the identification of optimum levels of design and process parameters at a moderate cost. EPA continues to fund research and development activity aimed at identifying new variants of the process and efforts toward improving the existing processes, including the use of limestone for regeneration.

This year is a major milestone in the application of dual alkali technology to full-scale utility systems in the U.S. Three full-scale systems ranging in size from 250 to 575 MW are scheduled to come onstream during 1979. The owners of these systems, their size, and the vendors providing them are: (1) Central Illinois Public Service Co., Newton No. 1, 575 MW unit, Envirotech; (2) Louisville Gas & Electric Co., Cane Run No. 6, 277 MW unit, Combustion Equipment Associates/Arthur D. Little, Inc.; and (3) Southern Indiana Gas & Electric Co., A.B. Brown No. 1, 250 MW unit, FMC. All of these systems are of the concentrated, sodium-based dual alkali type and they service boilers firing high-sulfur coal.

This paper presents a brief description, the design bases, and a summary of vendor guarantees and reported costs for these systems. In addition, it presents the fundamentals of dual alkali technology, reviews the status of commercial dual alkali systems, and briefly describes some of the non-sodium sulfite based dual alkali processes.

NOTES

1. Company Names and Products.

The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Consistency of Information.

The information presented was obtained from a variety of sources (sometimes by telephone conversation) including system vendors, users, EPA trip reports, and other technical reports. As such, consistency of information on a particular system and between the several systems discussed may be lacking. The information presented is basically that which was voluntarily submitted by developers and users with some interpretation by the author. The order of presentation of information or the amount of information presented for any one system should not be construed to favor or disfavor that particular system.

3. Units of Measure

EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL-RTP provides conversion factors for the non-metric units. Generally, this paper uses British units of measure.

The following equivalents can be used for conversion to the Metric system:

<u>British</u>	<u>Metric</u>
5/9 (°F-32)	°C
1 ft	0.3048 m
1 ft ²	0.0929 m ²
1 ft ³	0.0283 m ³
1 grain	0.0648 gram
1 in.	2.54 cm
1 in. ²	6.452 cm ²
1 in. ³	16.39 cm ³
1 lb (avoir.)	0.4536 kg
1 ton (long)	1.0160 m tons
1 ton (short)	0.9072 m tons
1 gal.	3.7853 liters

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<u>Individual</u>	<u>Company</u>
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Lee, G.C.Y.	Bechtel National, Inc.
Ramirez, A.A.	Food Machinery Corporation (FMC)
Van Ness, R.P.	Louisville Gas & Electric Co.
Wagner, N.P.	Southern Indiana Gas & Electric Co.

SUMMARY OF UTILITY DUAL ALKALI SYSTEMS

Section 1

INTRODUCTION AND BACKGROUND

1.1 INTRODUCTION

At the 1976 EPA Symposium on Flue Gas Desulfurization (FGD), held in New Orleans, it was reported that the EPA planned to co-fund a full-scale utility boiler dual alkali (D/A) demonstration program and that three viable proposals for installation of a full-scale, dual alkali, FGD system on coal-fired utility boilers had been received. The successful utility bidder was the Louisville Gas & Electric Co. (LG&E). LG&E had contracted to install the Combustion Equipment Associates/A.D. Little (CEA/ADL) designed D/A FGD unit at LG&E's Cane Run No. 6, 277 MW boiler. The demonstration monitoring program on the Cane Run FGD system is scheduled to begin in April 1979.

This is a memorable year for application of dual alkali technology to utility boilers in the U.S. because three full-scale systems, ranging in size from 250 to 575 MW, are scheduled for startup this year. The owners of these systems, their size, and the vendors providing them are: (1) Central Illinois Public Service Co., Newton No. 1, 575 MW unit, Envirotech; (2) Louisville Gas & Electric Co., Cane Run No. 6, 277 MW unit, CEA/ADL; and (3) Southern Indiana Gas & Electric Co., A.B. Brown No. 1, 250 MW unit, FMC. All of these systems are of the concentrated, sodium based dual alkali type and they service boilers firing high-sulfur coal.

This paper presents a brief description, the design bases, and a summary of vendor guarantees and reported costs for these systems. In addition it presents the fundamentals of dual alkali technology, reviews the status of commercial dual alkali systems, and briefly describes some of the non-sodium sulfite based dual alkali processes.

Before describing the three full-scale utility systems, a review of the technological background is warranted. Like any specialized technology, a host of terms for and variations of the sodium-sulfite based dual alkali processes have evolved. A detailed discussion of terms and the significant process, design and cost considerations is included in Appendices A and B. However, a brief qualitative discussion is given below.

1.2 BACKGROUND

"Double alkali" or "dual alkali" processes are characterized as non-recovery SO_2 abatement processes which involve aqueous alkali scrubbing of sulfur oxides from the flue gas, followed by regeneration of the scrubbing solution with lime or limestone to precipitate the sulfite/sulfate reaction waste. These processes are also referred to as "indirect" lime/limestone processes.

Continued developmental activity has resulted in several distinct process variations. However, common to each process is the separation of the absorp-

tion and the regeneration cycles. This separation permits the use of clear absorption solutions (i.e., no slurry in the scrubber) and results in the following advantages relative to the direct lime/limestone processes.

- The scaling, plugging, and erosion potentials within the scrubbing loop (i.e., absorption step) are greatly reduced.
- The SO₂ absorption efficiency is increased since the important rate limiting lime/limestone dissolution is not required in the scrubber.
- The higher SO₂ removal efficiency results in reduced absorbent liquor circulation and simpler scrubber requirement as compared to slurry systems.
- The production of waste solids is reduced by the higher utilization of lime or limestone in the regeneration process.

The typical dual alkali process is illustrated in Figure 1-1. It comprises the following basic operations:

- Absorption: reaction of sulfur oxides with an aqueous alkaline absorbent to form soluble sulfite and bisulfite ions in solution in the scrubber.
- Regeneration: treatment of a slip stream of the recirculating absorbent solution with lime or limestone to precipitate the insoluble calcium sulfite/sulfate waste product and increase the alkalinity of the absorbent. The regeneration process is spatially separated from the scrubbing operation.
- Dewatering: separation of the calcium-based precipitate from the absorbent liquor and recovery of alkali liquor.
- Softening: lowering the dissolved calcium ion concentration in the regenerated solution (i.e., subsaturating it with respect to gypsum) to reduce scaling potential in the scrubber. Sodium carbonate is generally used to precipitate calcium ion as calcium carbonate in dilute dual alkali systems.

Each dual alkali process is characterized by a specific cation (Na⁺, NH₄⁺, Mg⁺⁺, Al⁺⁺⁺), associated with the absorbent base, an alkaline earth regenerant (lime, limestone), and a solution strength at which the active absorbent can be regenerated (a discussion of dilute mode vs. concentrated mode D/A process is given in Appendix A). Regeneration is accomplished via the calcium sulfite/sulfate precipitation step.

With the exception of NH₄⁺, all of the above are non-volatile cations, thus enabling the use of simpler scrubbing systems. The loss of the volatile alkali, emission of visible plume of ammonia salts from the stack, and deposits of

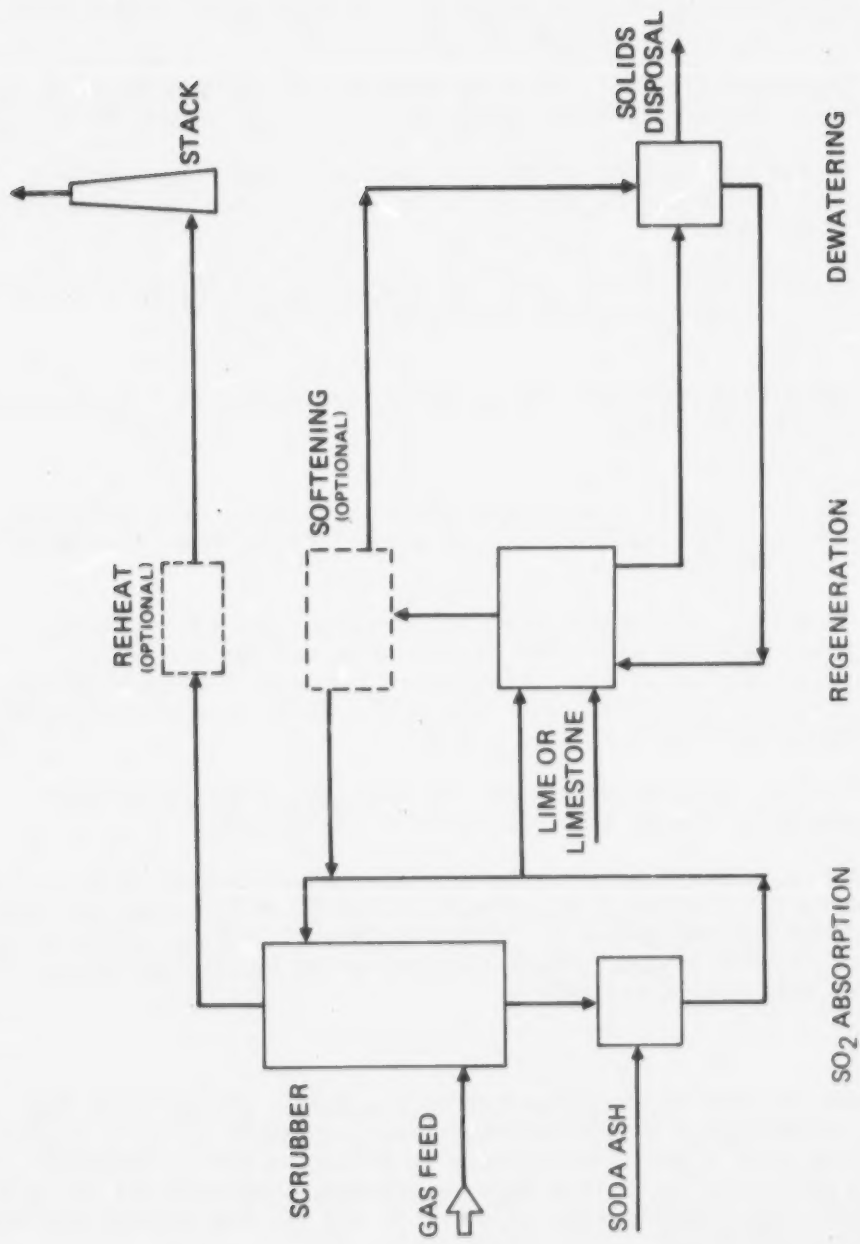


Figure 1-1 TYPICAL DUAL ALKALI PROCESS

ammonium compounds in the ducts and stacks have been reported as problems with ammonia scrubbers. The cost involved in the solution of these problems has retarded the development of the ammonia-based dual alkali scrubbing technology.

A status summary of operating and planned full-scale dual alkali systems in U.S. and Japan is given in Tables 1-1 and 1-2, respectively. The tables show a total of approximately 7,250 MW equivalent operating and planned dual alkali systems in industrial and utility applications in the U.S. and Japan representing 73 applications of this technology. Of these, approximately 5,775 MW equivalent, representing 65 applications are operational. In the U.S. eight operational units approximately equivalent to 252 MW are listed. As a standard for comparison, the latest PEDCo survey⁽²⁸⁾ of utility FGD systems in the U.S. indicates a total of 59,469 MW representing 139 units of operating and planned FGD systems. Of these, 14,480 MW representing 40 units are operational.

Table 1-1
SUMMARY OF SIGNIFICANT OPERATING AND PLANNED
FULL-SCALE DUAL ALKALI SYSTEMS IN THE U.S. (13, 28, 31)

PROCESS DEVELOPER	ABSORBENT, PRECIPITANT	ACTIVE ALKALI	USER	PLANT SITE	HW	TYPE OF PLANT ^(c)	YEAR OF COMPLETION ^(a)
Food Machinery Corporation (FMC)	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	FMC	Modesto, California	10 (Gas Rate) 30 (Regen.)	Reduction kiln ^(R)	1971
General Motors Corp.	$\text{NaOH}/\text{Na}_2\text{SO}_3$, $\text{Ca}(\text{OH})_2$	Dilute	General Motors Corporation	Parma, Ohio	32 (Gas Rate) 40 (Regen.)	Industrial boiler ^(R)	1974
Zurn Industries	$\text{NaOH}/\text{Na}_2\text{SO}_3$, $\text{Ca}(\text{OH})_2$	Dilute	Caterpillar Tractor Co.	Joliet, Illinois	20-30	Industrial boiler ^(R)	1974
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Firesome Tire & Rubber Co.	Pottstown, Pa.	3	Demonstration ^(R)	1975
Combustion Equipment Associates (CEA)/Arthur D. Little (ADL)	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Gulf Power Co.	Sneads, Florida	20	Utility boiler ^(R)	1975
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Caterpillar Tractor Co.	Mossville, Ill.	50	2 Industrial boilers ^(1M,1R)	1975
Zurn Industries	$\text{NaOH}/\text{Na}_2\text{SO}_3$, $\text{Ca}(\text{OH})_2$	Dilute	Caterpillar Tractor Co.	Morton, Illinois	12	Industrial boiler ^(R)	1978
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Caterpillar Tractor Co.	East Peoria, Ill.	100	4 Industrial boilers ^(2M,2R)	1978
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Caterpillar Tractor Co.	Mapleton, Ill.	140	Industrial boiler ^(M)	(1979)
Envirotech	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Central Illinois Public Service Co.	Newton, Illinois	575	Utility boiler ^(M)	(1979)
CEA/ADL	Na_2SO_3 , Carbide Lime	Concentrated	Louisville Gas & Electric Co.	Louisville, KY	277	Utility boiler ^(R)	(1979)
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Southern Indiana Gas & Electric Co.	Evansville, Ind.	250	Utility boiler ^(M)	(1979)
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Arco/Polymers, Inc.	Monaca, Pa.	100	Industrial boiler ^(R)	(1980)
FMC	Na_2SO_3 , $\text{Ca}(\text{OH})_2$	Concentrated	Chanslor Western Oil & Dev. Co.	Bakersfield, Ca	25	Industrial boiler ^(M)	(1979)
(b)	n.a.	n.a.	Dupont Inc.	Georgia, Al.	100	Industrial boiler ^(M)	(1987)
(b)	n.a.	n.a.	Grisson Air Force Base	Bunker Hill, Ind.	12	3 Industrial boilers ^(1M,2R)	(1979)

a) Dates in parentheses are projected start-up dates.

b) Vendor not selected.

c) M = New; R = Retrofit

Table 1-2
SUMMARY OF SIGNIFICANT OPERATING
FULL-SCALE DUAL ALKALI SYSTEMS IN JAPAN (1)

Process Developer	Absorbent, Percipitant	User	Plant site	Capacity (1,000 Nm ³ /hr)	Capacity (M)	Source of Gas	Inlet SO ₂ (ppm)	Year of Completion
Showa Denko	Na ₂ SO ₃ , CaCO ₃	Showa Denko Kanagafuchi	Chiba	580	178	Industrial boiler	1,500	1973
		Showa Denko Kawasaki	Tokasago	300	107	Industrial boiler	1,500	1974
Showa Denko-Ebara	Na ₂ SO ₃ , CaO	Wippon Hining	Saganoseki	200	71	Industrial boiler	1,400	1974
		Yokohama Rubber	Hiratsuka	120		H ₂ SO ₄ plant		1973
		Mitsubishi Oils	Isogo	105	38	Industrial boiler		1974
		Poly Plastics	Fuji	212	76	Industrial boiler		1974
		Ajinomoto	Yokkaichi	82	29	Industrial boiler		1974
		Kyowa Pet. Chem.	Yokkaichi	150	54	Industrial boiler		1974
		Japan Food	Yokkaichi	100	36	Industrial boiler		1975
		Yokohama Rubber	Mie	100	36	Industrial boiler		1975
		Asia Oil	Yokohama	243	87	Industrial boiler		1975
		Nippon	Keihin	150		Sintering plant	400	1975
		Kinura Utility	Nagoya	185	66	Industrial boiler		1972
		Daisowa Paper	Fuji	264	94	Industrial boiler		1974
Nippon Kokan Tsukishima Kurabo Eng.	(NH ₄) ₂ SO ₃ , CaO	Kuray	Tamashima	100	36	Industrial boiler	1,500	1975
		Daicel	Aoshi	163	58	Industrial boiler	1,300	1975
		Bridgestone Tire	Tosu	60	21	Industrial boiler		1975
		Bridgestone Tire	Tochigi	80	28	Industrial boiler		1975
Dowa Mining	Al ₂ (SO ₄) ₃ , CaCO ₃	Jujo Paper	Ishinomaki	200	71	Industrial boiler	1,200	1976
		Taenaka Mining	Mobara			Kiln	7,500	1972
		Dowa Mining	Okayama	150 x 2		H ₂ SO ₄ plant	650	1974
		Naikai Engyo	Okayama	70	25	Industrial boiler	1,500	1976
		Yahagi Iron	Nagoya	50		Sintering plant	2,500	1976
		Nihon Seiko	Nakase	30		Sintering plant	5,000	1976
		Kowa Seiko	Tobata	72		H ₂ SO ₄ plant	750	1978
		Mitsubishi Metal		140		Smelting furnace	4,000	1978
		Kureha Chemical	Nishiki	5	2	Industrial boiler	1,500	1975
		Kobe Steel	Wagasaki	175 x 2		Sintering plant	500	1976
		Kobe Steel	Kobe	350 x 2		Sintering plant	500	1976
		Osaka Steel	Osaka	375		Sintering plant	500	1976
Kureha Chemical Kobe Steel	CH ₃ COONa CaCl ₂ , CaO	Kakogawa Steel	Kakogawa	1,000		Sintering plant	500	1978
		Kobe Steel	Okazaki	200	71	Industrial boiler	1,600	1975
		Unitika	Saidaiji	300	107	Industrial boiler	1,400	1975
		Hippon Exlan	Shinsendai	420	150	Utility boiler	420	1974
		Tohoku Electric	Sakaide	1,260	450	Utility boiler	1,500	1975
		Shikoku Electric	Anan	1,260	450	Utility boiler	1,500	1975
		Kyushu Electric	Buzen	730	261	Utility boiler	1,500	1977
		Tohoku Electric	Akita	1,050	375	Utility boiler	1,500	1977
Kawasaki H. I.	MgO, CaCO ₃ MgO, CaO							
Kureha-Kawasaki	Na ₂ SO ₃ , CaCO ₃							

Section 2

DESIGN CRITERIA AND CURRENTLY OFFERED PROCESS GUARANTEES

Based on the performance and reliability demonstrated in various dual alkali pilot and prototype plants in the U.S. and Japan, and in an increasing number of full-scale applications in Japan, several general design and performance criteria are of interest. Principal among them are: (1) SO₂ removal performance, (2) particulate matter removal performance, (3) sodium consumption, (4) calcium consumption, (5) energy consumption, (6) waste solids quality, and (7) system reliability.

2.1 DESIGN CRITERIA

SO₂ Removal

A commercial dual alkali system must remove the desired quantity of SO₂ to allow compliance with the applicable Federal and/or local standards. With the present state-of-the-art, it is reasonable to expect long-term average SO₂ removal capability on the order of 95% with moderate, 10-15 gal./10³ acf, liquid-to-gas ratios, and high (up to 4,000 ppm) SO₂ inlet concentration.

Particulate Matter Removal

The scrubbed gas from the dual alkali flue gas desulfurization (FGD) unit should not contain particulate matter in excess of applicable standards. In some cases, this could require particulate matter removal by the FGD unit; in others it could merely imply no net addition of particulate matter to the gas stream by the FGD unit.

Sodium Consumption

Sodium consumption is an important performance criterion for a dual alkali system not so much from the viewpoint of economics but rather from its potential for secondary pollution. Sodium consumption is only a minor factor in the operating cost of a system, representing about 2% of the annual operating cost. Thus, even if soda makeup to a system were to increase by 100% over its design value, operating cost will be increased by a factor of less than 1.02. On the other hand, the environmental consequences of higher sodium consumption may be significant if the sodium is leached from the waste product to the environment.

A logical way to measure sodium consumption is in moles of Na consumed per mole of sulfur removed by the system. A value of 0.05 moles of Na makeup per mole of sulfur removed (equivalent to 0.025 moles of Na₂CO₃ make-up per mole of sulfur removed) appears to be a reasonable design target based on present U.S. technology. This target is achievable in concentrated dual alkali systems burning relatively high sulfur coal (over 3% sulfur) and having a relatively low oxygen content in flue gas. In Japan, sodium makeup is reportedly as low as 0.02 moles Na/mole sulfur removed for some systems.⁽¹³⁾

Calcium Consumption

A logical way to specify calcium consumption is as calcium stoichiometry; i.e., moles of calcium added per mole of sulfur removed (or collected). A calcium consumption of 0.98 to 1.0 appears to be a reasonable design target for concentrated dual alkali systems (values less than 1.0 are possible due to the alkali added with sodium makeup to the system).

Energy Consumption

Design targets for energy consumption can be in the range of 1-2% of power plant output without reheat, or under 4% even with 50 F° of reheat. These figures are based on a system which has a scrubber pressure drop in the range of 6-8 in. H₂O and a scrubber system L/G ratio of about 10 gal./10³ acf. This assumes that the power plant is equipped with some means of efficient particulate collection upstream of the FGD unit (e.g., an electrostatic precipitator).

Waste Solids Quality

The dual alkali process must produce easily handled, transportable, environmentally acceptable waste. Present state of technology indicates that a filter cake containing a minimum of 65 wt. % solids and a low level of soluble salts, such as Na₂SO₃ and NaCl, is a reasonable expectation. Systems currently being built in the U.S. (concentrated mode) will produce a waste solid containing primarily calcium sulfite with some amount of sulfate in the crystal. The Japanese systems are primarily designed to produce saleable gypsum.

System Reliability

The FGD system should have a high availability. Based on existing vendor guarantees availability of 90% for 1 year, as defined by the Edison Electric Institute for power plant equipment, may be a reasonable target.

2.2 CURRENTLY OFFERED PROCESS GUARANTEES

As examples of the currently offered process guarantees, those for the three, full-scale, utility dual alkali systems in the U.S. are summarized in Table 2-1. Because of site-specific factors, the reader is cautioned against any interfacility comparison or evaluation in light of the above-noted design criteria.

Table 2-1
FGD SYSTEM GUARANTEES (3)(22)(32)

	CIPSCO ^(a) Vendor: Envirotech	LG&E Vendor: CEA/ADL	SIEGECO Vendor: FMC
Availability	90% for 70% load factor for 30 year life span	90% for a 1-year operating period	95% for 30- and 60-day consecutive test runs
SO ₂ Removal	90% or outlet SO ₂ less than 200 ppm whichever is greater	200 ppm in scrubber outlet or 95% removal if sulfur content of the coal is greater than 5 wt. %	1.2 lb/10 ⁶ Btu (85% for 4.5 wt. % sulfur coal)
Particulate Matter Removal lb/10 ⁶ Btu	≤ 0.10 ^(b)	≤ 0.10 and no net addition of particulate	≤ 0.10 and no net addition of particulate
HCl Removal	90%	Not applicable	Not applicable
Sodium Consumption moles Na ₂ CO ₃ /mole SO ₂	0.024 (per mole SO ₂) and 0.023 (per mole of acid gas)	0.045 when maximum coal chloride level is 0.06% 0.5 moles additional for each mole of chloride in the coal above the 0.06% level	0.03 plus Na lost with chloride
Calcium Consumption moles Ca/mole SO ₂ removed	1.10	Maximum 1.05	Approximately 1.0
Energy Consumption	None	1.2% of peak operating rate (300 MW)	Less than 1% of operating rate
Solids Quality	None	Minimum 55 wt. % insoluble solids	Minimum 55 wt. % solids

a) Figures are not guarantees but rather expected design values.

b) FGD system designed for possible ESP upsets.

Section 3

FULL-SCALE UTILITY DUAL ALKALI SYSTEMS IN THE U.S.

This year (1979) three U.S. full-scale, utility boiler dual alkali systems, ranging in size from 250 to 575 MW, are scheduled for startup. The owners of these systems, their size, and the vendors providing them are: (1) Central Illinois Public Service Co., Newton No. 1, 575 MW unit, Envirotech; (2) Louisville Gas & Electric Co., Cane Run No. 6, 277 MW unit, CEA/ADL; and (3) Southern Indiana Gas & Electric Co., A.B. Brown No. 1, 250 MW unit, FMC. All of these systems are of the concentrated sodium based dual alkali type and they service boilers firing high sulfur coal.

Brief descriptions of the these full-scale dual alkali facilities, their design bases, and reported costs are presented in this section.

3.1 FACILITY DESCRIPTIONS

Central Illinois Public Service Co. (CIPSCO)⁽³⁾(10)

Boiler

Newton Station Unit No. 1, Newton, Illinois, utilizes a pulverized coal-fired boiler manufactured by Combustion Engineering Inc. It is designed for a maximum steam generating capacity of 4,158,619 lb/hr at 2620 psig and 1005°F at the steam outlet connection. Design excess air in the boiler is 22% with flue gas flow to the scrubbing system totalling 6,615,000 lb/hr. The boiler design includes two parallel induced draft fans with discharge into a common plenum which can feed flue gas to the FGD system or to the stack. The boiler is tangentially fired with coal from bowl-type pulverizing mills utilizing high-sulfur Illinois bituminous coals from several local sources.

FGD Facility

Process equipment of interest, in the "duplex-type" dual alkali FGD system, built to serve Newton No. 1, includes:

- Four booster fans.
- Four precoolers spray towers, including mist eliminators.
- Four mobile-ball gas scrubbers, including mist eliminators.
- Three spent absorbent causticizers (reactors).
- One precoolers effluent neutralizer.

- One reactor/clarifier.
- Two 100-ft diameter thickeners, for the dual alkali solids concentration system, with concrete bottoms and access to bottom discharge cone through a tunnel.
- One 50-ft diameter thickener, for the precool loop, constructed of coated steel.
- Three horizontal Eimco-Extractor filters for dewatering and multi-stage washing of thickener underflow.

Figures 3-1 and 3-2 are the process flow diagram and a general view of the FGD system, respectively. An overview of the absorption, regeneration, and dewatering sections is given below.

Absorption Section

Precooler/HCl Absorber: Because of design criteria requiring the use of Total Illinois coal, the system is designed to accommodate these high-chloride coals. The presence of chlorides in the coal may require a preconditioning step before the flue gas is treated for SO₂ removal in the main scrubbing loop. This preconditioning of the flue gas in the precooler spray towers also minimizes soda ash consumption.

The precooler spray towers treat and cool the flue gas. The towers are designed to remove 90% of the chloride in the form of HCl. To minimize any carryover of mist or particulate matter, vertical mist eliminators are installed in the horizontal precooler tower outlet ducts (liquid drains perpendicular to the gas flow). These mist eliminators control the acid mist within the precooler loop and the level of dissolved solids (chlorides) in the scrubbing liquor to reduce sodium consumption and minimize operational problems.

The spent liquor from the precooler is sent to the fly ash thickener for clarification and prevention of significant concentrations of abrasive slurry solids in the recirculating liquor. Also, to maintain steady sodium level, pH, and solids level in the precooler loop, the clarifier underflow is dewatered and treated with lime and sodium carbonate. The regenerated liquor is returned to the precooler loop.

SO₂ Scrubber: SO₂ emissions are reduced to less than 200 ppm utilizing a countercurrent, two-tray-stage mobile-ball bed scrubber operating at 8.3 ft/sec design gas velocity with a L/G of less than 10 gal./10³ acf). A mist eliminator mounted vertically in the horizontal duct (liquid drains perpendicular to the gas flow) controls sodium liquor carryover. The eliminator vanes are intermittently washed with service water, which is collected and returned to the precooler loop to maintain a favorable water balance.

In the initial operation of the unit, two types of reheat systems are used (half the gas volume for each):

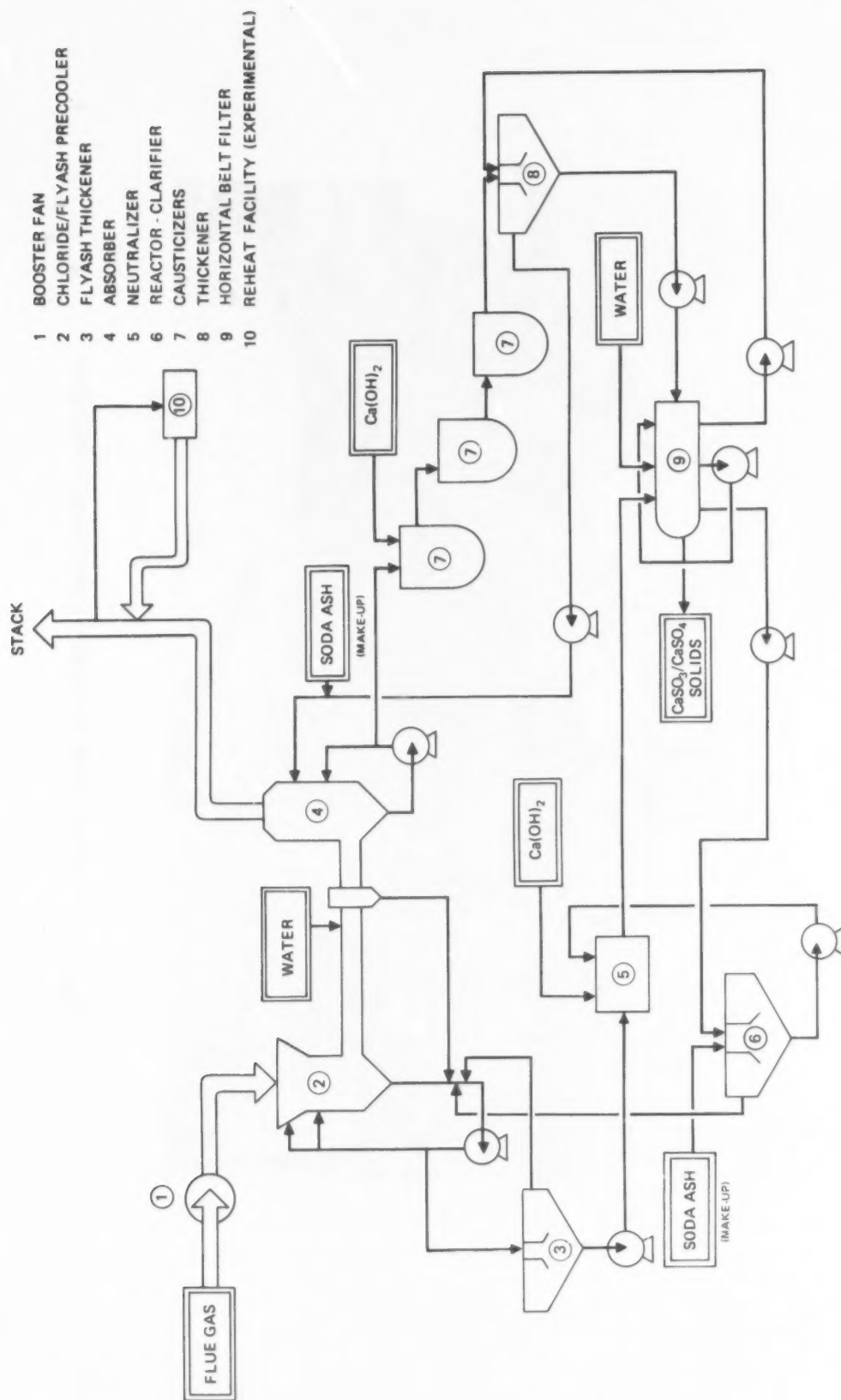


Figure 3-1 ENVIROTECH FLUE GAS DESULFURIZATION SYSTEM ON CIPSCO'S NEWTON NO. 1 UNIT

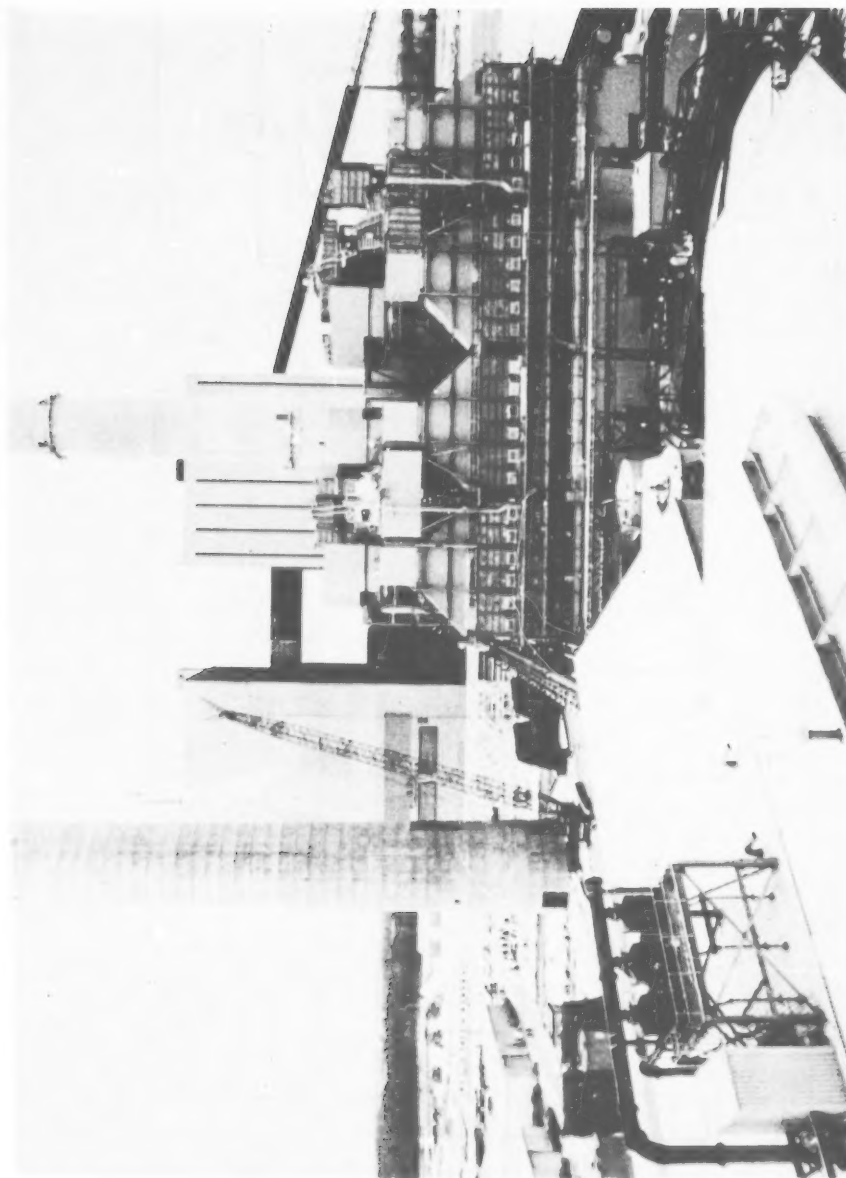


Figure 3-2 GENERAL VIEW OF CIPSCO NEWTON NO.1 FGD UNIT

1. Recycle Reheat

2. Bypass Reheat

With the recycle reheat mode of operation, the reheat is accomplished by blending a side stream of steam-heated flue gas with the saturated flue gas exiting the scrubber trains. The blending occurs across a high-alloy perforated-plate diffuser section in the main duct. The steam heated flue gas is maintained above the saturation point during normal operation by recycling and preheating gas on the downstream side of the diffuser. The gas entering the reheater will be above the saturation point to reduce corrosion potential. The gas in the recycle reheat section during normal operation will range between 156°F and 300°F, as compared to the saturated gas temperature of 131°F. The system is designed to achieve an exit gas temperature of 156°F to the stack.

In the Bypass Reheat method, a portion of the unscrubbed flue gas is bypassed around the scrubber and mixed with the cleaned, saturated, scrubbed gases exiting the scrubber. A diffuser section, as described above, is used to disperse and mix the two gas streams. This system is designed to achieve a maximum of 43°F reheat because of the limitation on allowed SO₂ and particulate emission levels. The quantity of bypass gas is controlled by damper control to realize the desired degree of reheat. In addition, the gas is also monitored in regard to opacity and SO₂ level as an additional means of control.

The overall design philosophy in regard to the reheat systems was to provide sufficient flexibility to allow either type of reheat system to be installed later if one of the systems proved to be more effective. Space has been reserved to achieve up to 50 F° of reheat, based on space requirements for the larger system.

Regeneration Section

A portion of the spent liquor from the absorber is pumped to the first stage of a three-stage causticizer (reactor) system, where slaked lime slurry is mixed with spent liquor.

The primary causticizer overflows by gravity into the secondary causticizer which in turn overflows into the tertiary causticizer. The slurry from the tertiary causticizer is pumped to the dewatering section. It is not certain whether one, two, or three reactor stages will be operated. If one is needed, there will be two spares. Likewise, if two are needed one will act as a spare.

Dewatering Section

After thickening of the sulfur oxides solids collected in the mobile-ball scrubber, the solids are dewatered and fresh-water washed in a top-loading horizontal-belt-type extractor filter. The unique geometry of this equipment permits counter-current multi-stage washing of the raw cake with limited quantities of service water. Over 80% of the entrained sodium in the vacuum filter cake is recovered. This sodium is recycled to the absorption loop for reuse. In addition, the liquid purge stream from the precoolers loop, containing collected chlorides, residual fly

ash, and trace elements, flows to a lime-neutralization tank and then is utilized as the wash medium for a final "cake-impregnation wash." Thus, high-chloride, low-pH liquor bypasses the SO₂ absorber and is discarded as surface moisture in the final waste cake.

Louisville Gas & Electric Co. (LG&E)⁽³²⁾

Boiler

Cane Run Station Unit No. 6, Louisville, Kentucky, utilizes a pulverized coal-fired boiler manufactured by Combustion Engineering Inc. It is designed for a maximum steam generating capacity of 1,854,217 lb/hr at 2600 psig and 1005°F at the steam outlet connection. Design flue gas flow to the scrubbing system totals 3,372,000 lb/hr. The boiler is tangentially fired with coal from bowl-type pulverizing mills utilizing, high-sulfur, midwest bituminous coals.

FGD Facility

Process equipment of interest, in the dual alkali FGD system for Cane Run No. 6, includes:

- Two booster fans.
- Two dual-tray absorbers.
- Two pairs of spent absorbent regeneration reactors.
- One 125-ft diameter thickener with concrete bottom and access to bottom discharge cone through a tunnel.
- Three rotary-drum vacuum filters with water wash.
- Two external combustion, scrubbed gas reheaters.

Figure 3-3 is a process flow diagram of the FGD System. Figures 3-4, 3-5 and 3-6 are photographs showing a general view of Cane Run Nos. 4, 5 and 6 FGD units, Unit 6 regeneration and dewatering areas, and Unit 6 vacuum filter, respectively. An overview of the absorption, regeneration, and dewatering sections is given below.

Absorption Section

The flue gas from the existing electrostatic precipitator (ESP) induced draft fan is forced by a booster fan into an absorber. There are two absorber modules, each equipped with a booster fan. A common duct connects the two inlet ducts to the booster fans.

The hot flue gas is adiabatically cooled and saturated by sprays of absorber solution directed at the underside of the bottom tray. These sprays keep the underside of the tray and the bottom of the absorber free of buildup of fly ash solids. The cooled gas then passes through two sieve trays, where SO₂ is removed, and

- 1 FORCED DRAFT FAN
- 2 SPRAY/TRAY ABSORBER
- 3 MIST ELIMINATOR
- 4 PRIMARY REACTOR
- 5 SECONDARY REACTOR
- 6 THICKENER
- 7 FEED FORWARD TANK
- 8 VACUUM FILTER
- 9 FILTRATE SUMP
- 10 REHEAT FACILITY
- 11 MIX PLANT

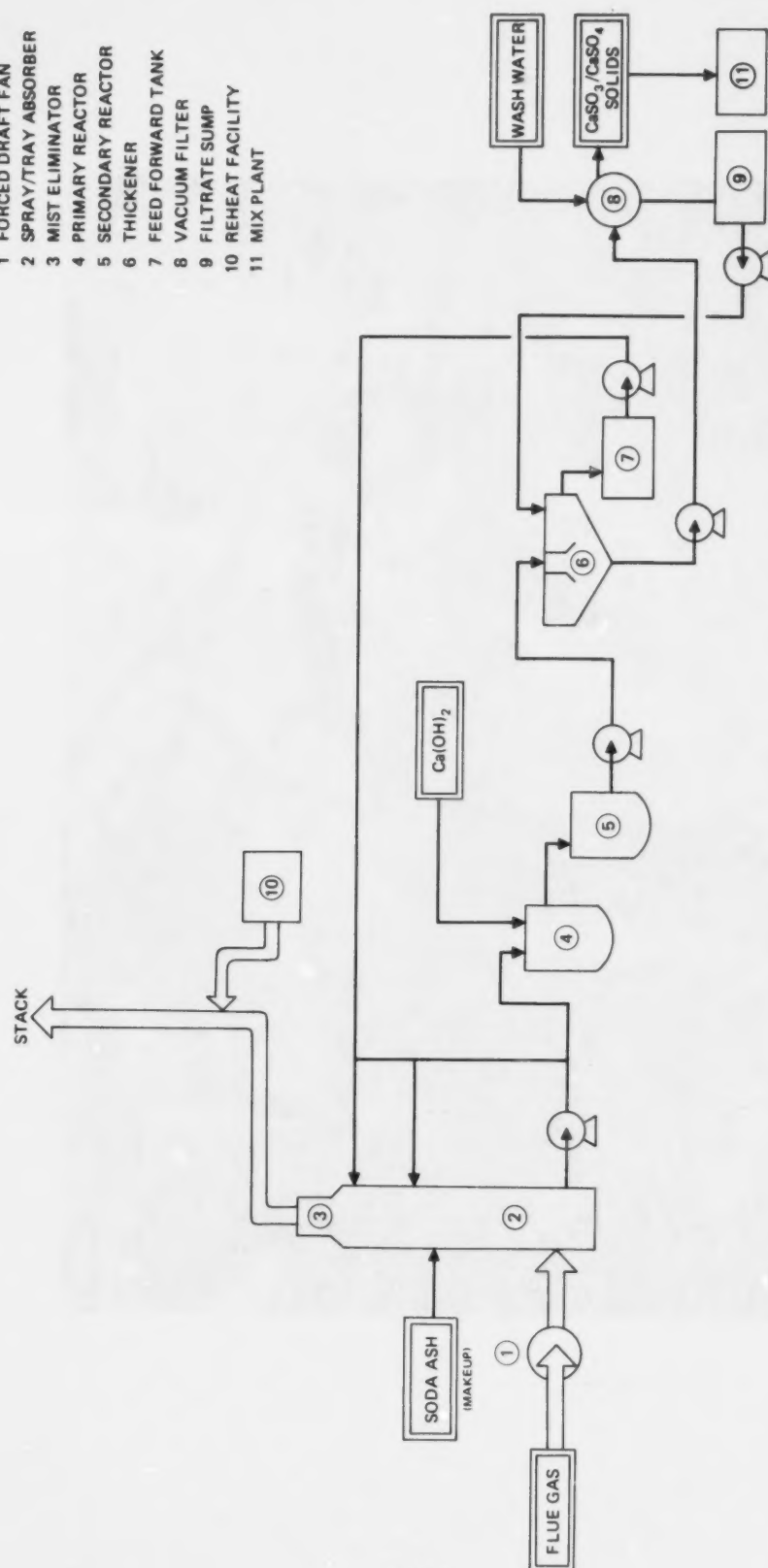


Figure 3-3 CEA/ADL FLUE GAS DESULFURIZATION SYSTEM ON LG & E'S CANE RUN NO.6 UNIT

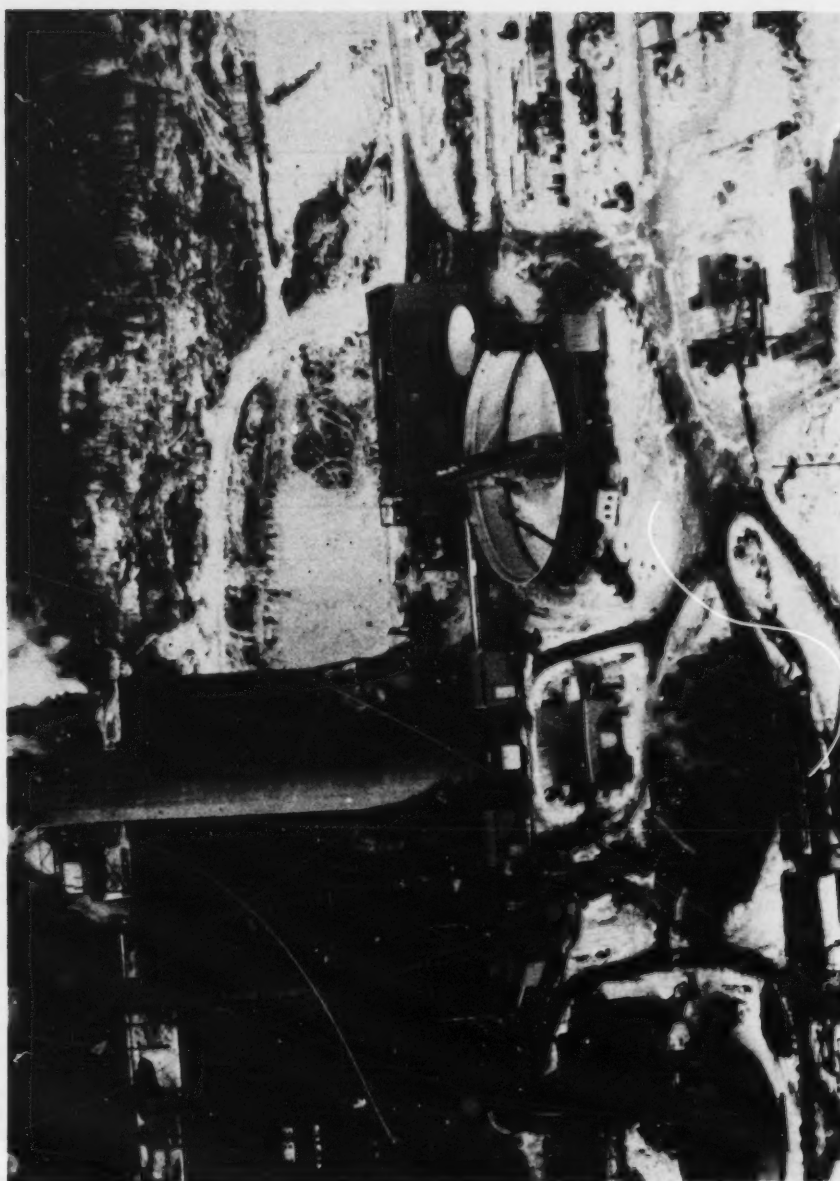


Figure 3-4 GENERAL VIEW OF LG&E CANE RUN NOS. 4, 5, AND 6 FGD UNITS

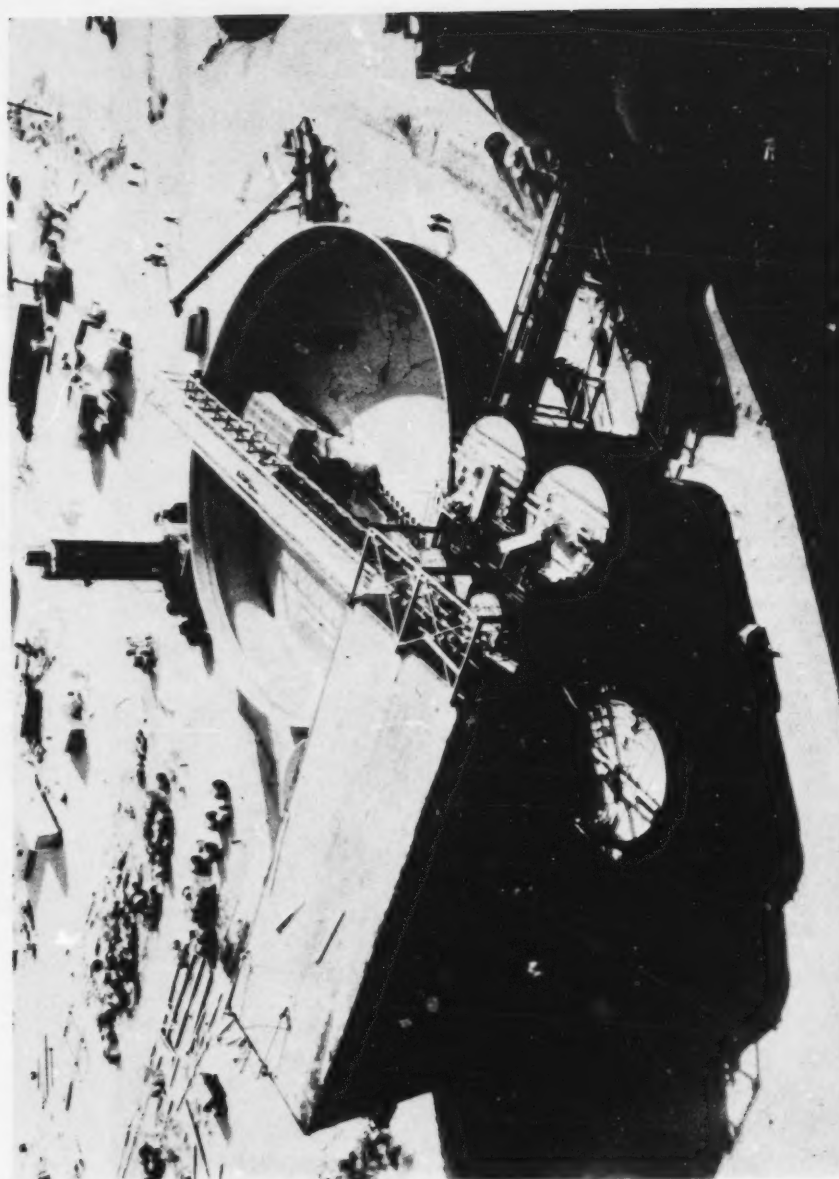


Figure 3-5 CANE RUN NO. 6 FGD UNIT REGENERATION AND DEWATERING AREAS

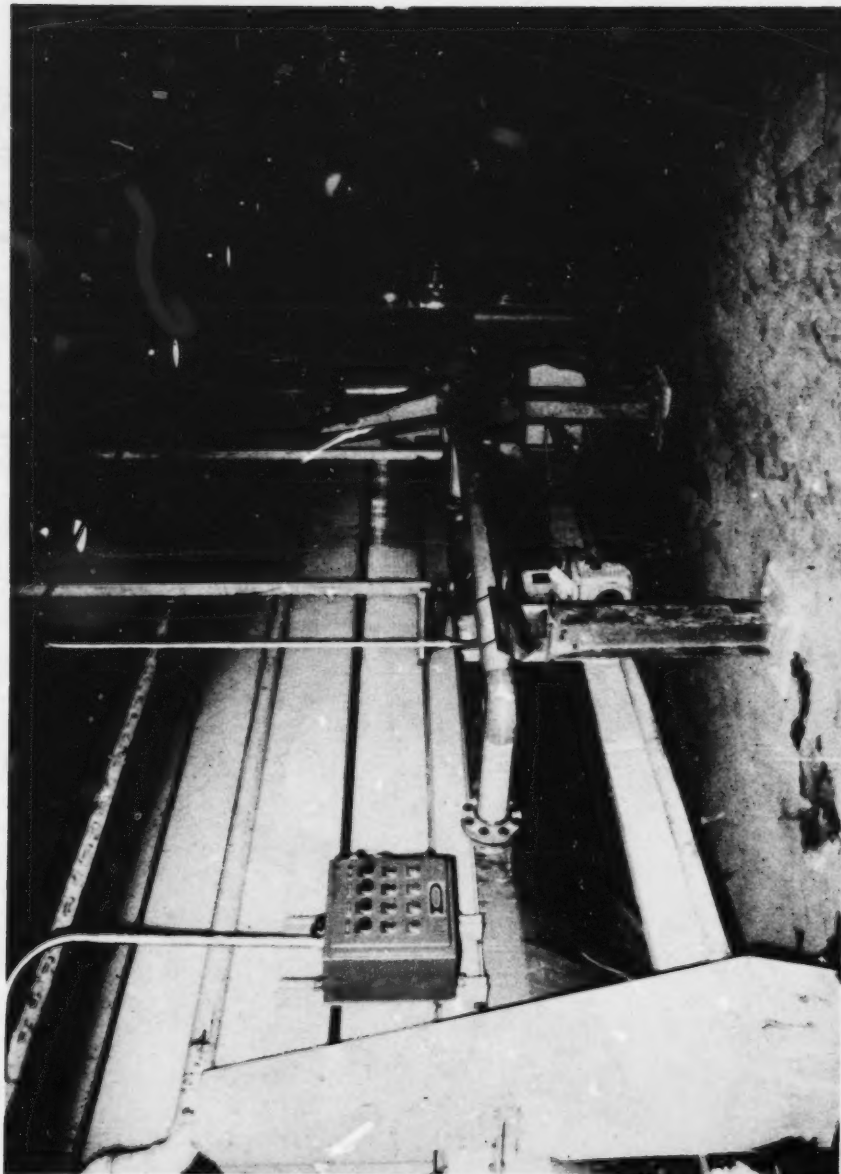


Figure 3-6 CANE RUN NO. 6 FGD UNIT VACUUM FILTER

leaves the absorber through a chevron type mist eliminator. Prior to entering the stack, the saturated gas from the mist eliminator is heated 50°F, to 175°F, by hot combustion gas from a grade-mounted reheater fired with No. 2 oil.

Each absorber is designed for 9.0 ft/sec gas velocity, a rate consistent with good mass transfer, low pressure drop, and minimal entrainment. Each absorber is sized to handle 60% of the design gas flow rate and the overall system can be turned down to 20% of the design flow rate by shutting down one absorber. At levels less than 50% of design capacity, the system can be operated with one absorber module by use of the common inlet duct.

For control of tray feed liquor pH, regenerated scrubbing liquor from the thickener hold tank is mixed in-line with absorber recycle liquor for each unit; the mixture is then fed to the top tray in each absorber. The absorber recycle liquor is used in the spray section below the trays. A bleedstream of the absorber recycle liquor is withdrawn and sent to the reactor system for regeneration. The bleed rate is controlled by the liquid level in the absorber. The feed forward rate of regenerated liquor from the thickener hold tank to the scrubber trays corresponds to an L/G of 4.0 gal./10³acf (saturated) at design conditions. The total recirculation rate for each absorber (sprays plus trays but excluding the feed forward regenerated liquor) corresponds to an L/G of 5.7 gal./10³acf. Thus, the total L/G is about 10 gal./10³acf.

Regeneration Section

The spent liquor (absorber bleed) is fed to the primary reactor of the two-stage reactor system along with slurried carbide lime. The primary reactor has a nominal liquor holdup time of 4.5 minutes at design flow. The primary reactor overflows by gravity into the secondary reactor, which has a nominal holdup of 40 minutes at design flow, where the reaction between lime and sodium salts is completed. The reaction product is a slurry containing 2-4% insoluble calcium salts and the regenerated sodium salt solution. The reactor slurry is pumped to the solid/liquid separation section. The pumping rate is controlled by the liquid level in the reactor.

Two reactor trains are provided, each train consisting of a primary reactor, a secondary reactor, and a reactor pump. At design conditions, both of the reactor trains would normally be in service, with each reactor train regenerating the spent liquor from the corresponding absorber. The reactor trains are identical, and each can be operated on liquor from either absorber or combined liquor from both the absorbers. For short-term, a reactor train may handle the total liquor from both the absorbers operating at design conditions. Thus, maintenance can be performed on one reactor train while the other is operating.

Dewatering Section

The reactor effluent streams are fed to the thickener. Clarified liquor overflows to the thickener hold tank from which the regenerated solution is pumped automatically to the absorbers to maintain the pH of the absorber liquor. The total volume in the system is maintained by controlling the liquid level in the thickener hold tank using process makeup water.

The thickener underflow slurry, controlled at about 25 wt.% solids, is pumped to the filter system where solids separation is completed. The filter cake is washed with fresh water to recover the sodium salts in the liquor. Combined filtrate and wash water are returned to the thickener.

There are three rotary drum vacuum filters, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (to obtain cake containing high solids content and low soluble salts content) it is desirable to operate the filters at fixed conditions (constant drum speed, submergence, wash ratio, etc.). Therefore, the cake rate is controlled by changing the number of filters in operation. The number of filters in operation is determined by the amount of solids accumulated in the thickener, which is reflected in the solids concentration in the underflow slurry. The density of the underflow slurry is measured and thickener hold tank liquor is added as required to maintain the percent solids in the underflow slurry at about 25%. The number of filters in operation is changed if the concentration of solids in the underflow slurry cannot be controlled using the dilution liquor.

Southern Indiana Gas & Electric Co. (SIGECO)⁽²²⁾

FGD Facility

Process equipment items of interest, in dual alkali system for Unit 1, include:

- Two booster fans.
- Two, three-stage disc contactors.
- One lime reactor.
- One, 100-ft diameter, thickener for dual alkali solids concentration.
- Three, rotary-drum, vacuum filters with water wash.

Figures 3-7 and 3-8 are the process flow diagram and a view of the FGD system absorber, respectively. A description of the absorption, regeneration, and de-watering sections is given below.

Absorption Section

The FGD system receives flue gas from two points on the discharge side of the electrostatic precipitator. The flue gas is drawn into two fans to increase the static pressure to the level required to force the gas through the absorbers and to the stack.

The SO₂ absorbers are low pressure drop, three-stage, counter-current disc contactors. Liquid/gas contacting is accomplished with fixed position flow diverters. The units are 30 ft-5 in. in diameter and have an overall height to the discharge of 70 ft. The top of the scrubber contains a chevron type mist eliminator to prevent liquid entrainment. This mist eliminator is washed approximately once per shift to remove any accumulation of salts and/or particulate matter.

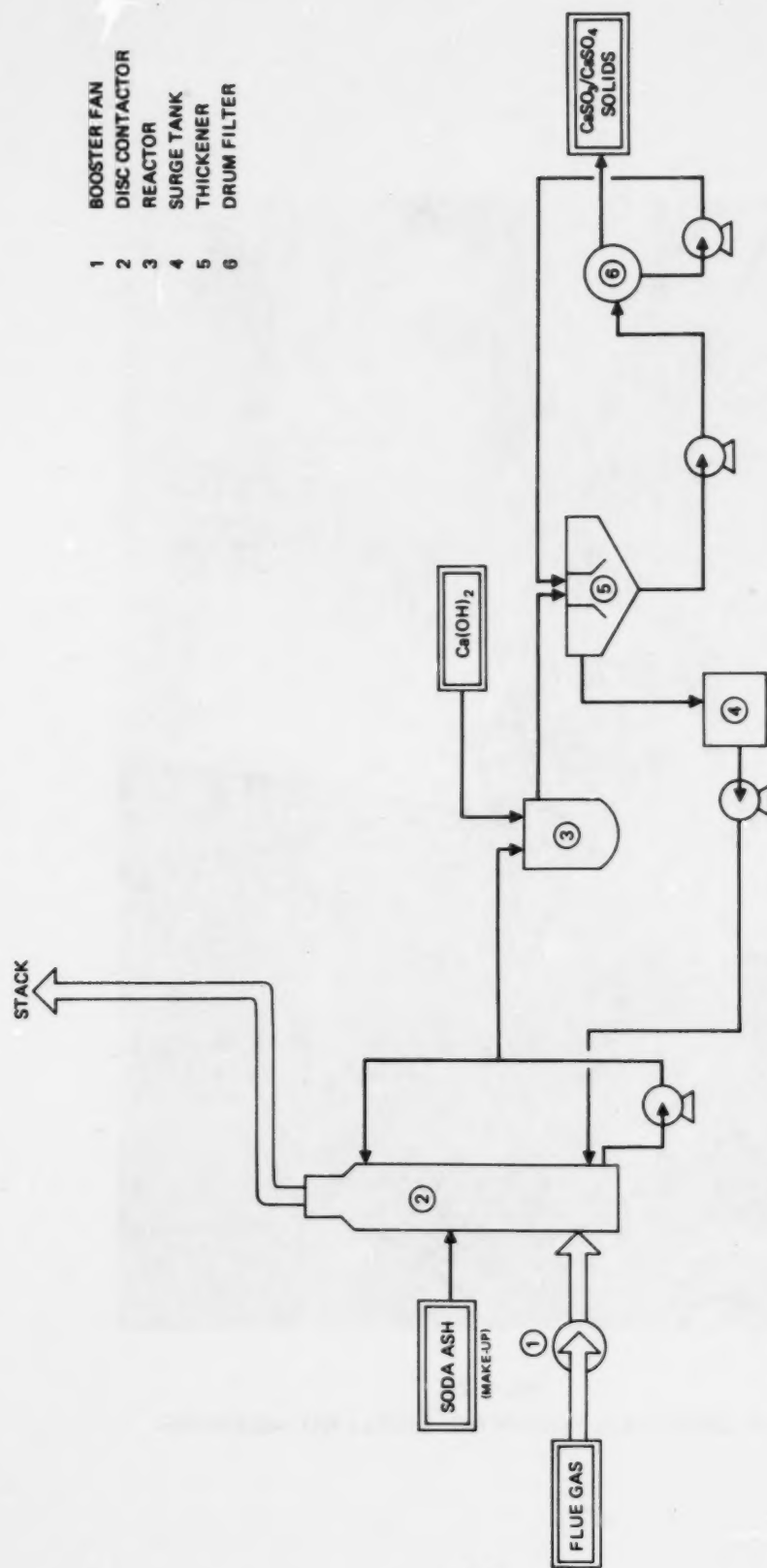


Figure 3-7 FMC FLUE GAS DESULFURIZATION SYSTEM ON SIGECO'S A.B. BROWN NO.1 UNIT

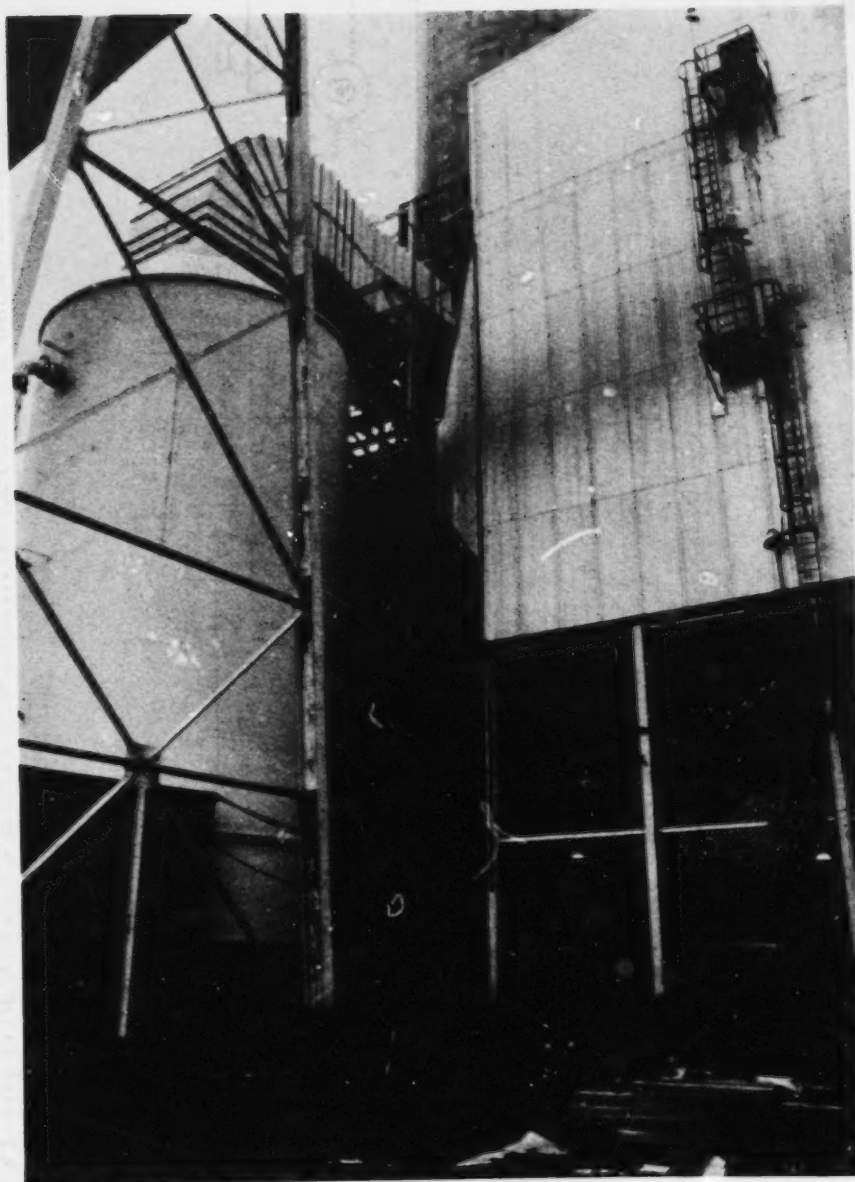


Figure 3-8
VIEW OF SIGECO A.B. BROWN NO. 1 FGD UNIT ABSORBER

Four manways are provided on each scrubber to give access to the mist eliminator, scrubber internals, and liquid sump. The bottom of the vessel is used as a reservoir for the scrubbing solution. During normal operation, the flue gas provides adequate heat to prevent freezing of the scrubbing solution. During system shutdown and subsequent startup, a steam sparger is provided to heat the solution. The bottom 8 ft of the contactor is insulated with fiberglass and covered with an aluminum jacket.

Two rubber-lined recirculation pumps circulate scrubbing liquor to the top stage of each scrubber. Each pump is capable of supplying 5200 gal./min. of solution. One pump is the primary operating pump; the second pump provides 100% spare capacity. The pumps are piped into the system with check valves to permit remote selection and startup without manual valve changes. A crossover starting system automatically starts the spare pump when flow fails. The flow of recirculated solution, not modulated for scrubbing load, is set for maximum gas flow.

Regeneration Section

Scrubbing solution to be regenerated is pumped to the lime reactor where it is neutralized with the lime slurry. The pH of the lime reactor is controlled by two separate pH control loops, one in operation and one spare.

The system has high and low pH alarms. A selector switch allows the operator to convert from one pH analyzer element to the other. The pH analyzer probes in this service are cleaned periodically. The regenerated slurry overflows from the lime reactor to the thickener where the calcium sulfite solids settle out of the regenerated solution. The rate of soda ash solution makeup to the scrubber modules controls the regenerated solution density. Density is an indication of the total solution concentration which must be maintained for optimum operation of the dual alkali process. The thickener underflow is pumped by diaphragm pumps to three rotary drum vacuum filters. The slurry flow rate to each filter is controlled by a low level switch. The regenerated scrubbing solution is collected in the regeneration surge tank and pumped back to the scrubbers based on pH demand.

Liquid level in the regeneration surge tank is controlled by addition of makeup water. The temperature of the surge tank is monitored and, in the event of low temperature, the tank is heated with steam supplied through a sparger.

Dewatering System

The underflow from the thickener is pumped to three rotary drum vacuum filters where the calcium sulfite precipitate is separated from the sodium sulfite solution. Cake removal is facilitated by low pressure air blown through the filter cloth just before the scrapers at the discharge. The filtrate is drawn into a filtrate receiver and is pumped back to the thickener. The vacuum is produced by a mechanical vacuum pump using water flushing. The water for this service is used only once.

3.2 DESIGN BASES

The design bases and design features for the three, utility dual alkali FGD system are presented in Tables 3-1 and 3-2, respectively. Typical analyses of the coals fired are summarized in Table 3-3.

Information in these tables leads to several general observations:

- All installations service boilers firing high-sulfur coal. Design coal sulfur range is 4.0 to 5.0 wt.%. Inlet SO₂ entering the absorbers ranges between 2600 and 3500 ppm.
- Design coal chloride content varies widely from 0.04 to 0.20 wt.%.
- Excess air including preheater leakage varies considerably as indicated by the 3.2 to 7.6% flue gas O₂ content range.
- Flue gas flow rate per scrubber module ranges between 395,000 and 522,000 acfm at 300°F.
- The number of regeneration reactors varies between designs, indicating differing design philosophies.
- All facilities use primary (thickener) and secondary (vacuum filters) dewatering devices. Sodium is recovered by wash schemes ranging from single- to multi-stage.
- Planned ultimate disposal of filter cake is to be accomplished, in all cases, by landfill. At the Newton No. 1 and Cane Run No. 6 plants, stabilization efforts may range from simple mixing with fly ash to mixing with fly ash and lime.

3.3 REPORTED COSTS

Before presenting the reported costs, a word of caution is in order. Because of many site-specific factors and/or different bases, quantitative evaluation of capital and operating costs for the various dual alkali processes for comparison with each other and with lime/limestone scrubbing is inadvisable.

Site-specific parameters that influence the capital cost of FGD systems include location and size of plant, new versus retrofit application, presence or absence of upstream fly ash removal facilities, fuel sulfur content, fuel chlorine content, degree of spare capacity, solid waste disposal method, and cost of real estate. The different cost factors will not apply equally to all processes. Each application requires separate assessment at conditions specified for a given installation to evaluate the cost ranking of the proposer's dual alkali systems.

Likewise, for operating costs, the impact of raw material (e.g., quick lime @ \$40/ton vs. carbide lime @ \$13.29/ton), utilities, and maintenance costs varies according to the type of process selected and applicable site-sensitive parameters. In addition, financial parameters, such as taxation and capital charges, vary with the location and accounting practice of the utility.

Table 3-1
FGD SYSTEM DESIGN BASES (3) (22) (32)

	CIPSCO NEWTON NO. 1	LG&E CANE RUN NO. 6	SIEGCO A.B. BROWN NO. 1
<u>Coal (Dry Basis):</u>			
Sulfur, wt. %	4.0	5.0	4.5
Chloride, wt. %	0.20	0.04	0.05
Heat Content, Btu/lb	10,900	11,000	13,010
<u>Inlet Gas:</u>			
Flow Rate (volumetric), acfm	2,163,480 ^(a)	1,065,000	790,036
Flow Rate (weight), lb/hr	6,615,000	3,372,000	2,415,764
Temperature, °F	327	300	290
SO ₂ , ppm	2590 ^(a)	3471	2810
O ₂ , %	7.7	5.7	3.2 ^(a)
Particulate, lb/10 ⁶ Btu	not available	0.10	0.10
<u>Outlet Gas:</u>			
SO ₂ , ppm	200	200	520
Particulate, lb/10 ⁶ Btu	0.10 ^(b)	0.10	0.10

(a) Calculated from other data.

(b) System designed to accommodate ESP upsets.

Table 3-2
FGD SYSTEM DESIGN FEATURES (3) (22) (32)

	CIPSCO VENDOR: ENVIROTECH	LG&E VENDOR: CEA/ADL	SIGECO VENDOR: FMC
FGD Unit Rating, Mw Process	575 Sodium/Calcium Hydroxide Dual Alkali/Concentrated ESP	277 Sodium/Carbide Lime Dual Alkali/Concentrated ESP	250 Sodium/Calcium Hydroxide Dual Alkali/Concentrated ESP
Fly Ash Collection No. of Modules Module Design	Four High velocity cocurrent spray tower, countercurrent two- tray-stage polysphere absorber Three reactors in series for four modules(a)	Two Two-stage tray tower absorber Two reactors in series per module Primary - 4.5 minutes Secondary - 40 minutes	Two Three-stage disc contactor One reactor per two modules
Regeneration Reactor Reactor Residence Time	No information	Three rotary drum vacuum filters	No information
Dewatering	Three horizontal belt filters	Fly ash stabilization/onsite landfill	Three rotary drum vacuum filters Transported to landfill without stabilization
Filter Cake Disposal	Fly ash stabilization/onsite landfill	Scrubbed gas reheat facility; None	
Additional Equipment	Chloride/fly ash precooler; fly ash thickener; neutralizer; reactor clarifier; and two ver- sions of experimental reheat facilities		

(a) In actual operation, one, two or three reactors may be used.

Table 3-3
TYPICAL COAL ANALYSES (3) (22) (32)
(Dry Basis)

<u>Species</u>	CIPSCO NEWTON NO. 1 Value, wt. %	LG&E CANE RUN NO. 6 Value, wt. %	SIGECO A.B. BROWN NO. 1 Value, wt. %
<u>ULTIMATE</u>			
Carbon	66.70	67.15	72.19
Hydrogen	4.92	4.72	5.01
Nitrogen	1.68	1.28	1.56
Chlorine	0.22	0.04	0.05
Sulfur	4.47	4.81	3.97
Ash	14.19	17.06	9.88
Oxygen	7.82	4.94	7.34
	100.00	100.00	100.00
<u>PROXIMATE</u>			
Moisture	10.50	8.95	11.34
Heat Content, Btu/lb	10,900	11,000	13,010

Actual capital and operating costs for full-scale, utility, dual alkali systems in the U.S. are as yet not available; construction of the first three, commercial-size systems is scheduled for completion in 1979 (see Table 1-1). Estimated costs for these applications are therefore based only on costs for prototype and small-scale commercial units, or on vendor and engineering estimates. Published cost data for the small-scale systems are fragmentary and have not been determined on equivalent bases. Generally, estimates by vendors tend to be low since they reflect vendor supplied portions of the project; a significant number of owner-incurred cost items are commonly omitted. Engineering estimates are usually based on specific applications and cannot be readily extrapolated for different conditions.

However, for informational purposes only, the capital and operating cost data are included herein. Tables 3-4 and 3-5 summarize those for the LG&E installation; Table 3-6 contains those for the SIGECO facility.

Table 3-4
ESTIMATED CAPITAL COSTS FOR THE DUAL ALKALI
FGD SYSTEM ON LG&E CANE RUN NO. 6 UNIT (32)

	<u>Dollars</u>
Total Materials Costs:	
Major Equipment Cost	7,037,000
Other Materials Cost ^a	2,525,000
Sludge Disposal Equipment	900,000
Additive Supply System ^b	700,000
Total Materials Cost	<u>11,162,000</u>
Erection: ^d	
Direct Labor (252,800 hrs. @ \$12/hr) ^c	3,034,000
Field Supervision	273,000
Total Erection Cost	<u>3,307,000</u>
Engineering Costs:	
System Supplier Engineering	1,323,000
Owner's Engineering Expenses	303,000
Owner's Consulting Engineer	852,000
Total Engineering Cost	<u>2,478,000</u>
Spare Parts, 2% of the Total Materials Cost	223,000
Working Capital	<u>200,000</u>
Total Capital Investment ^e	17,370,000
\$/kW (Based on 300 MW gross peak load)	57.9
(Based on 277 MW gross net load)	62.7

Basis: 300 Mw gross peak load existing coal fired boiler (277 Mw net peak load)
S in coal, 5.0%
SO₂ removal efficiency in scrubber, 94.2%
Stack gas reheat, 50% - direct oil fired reheater
Disposal of sludge after treatment to onsite pond
Project beginning mid 1976, ending late 1978, Avg. 1977 dollars
Necessary parts in storage and reasonable spare capacity

^a Sludge disposal equipment cost (\$900,000) is shown separately. The \$900,000 is the 3/7 portion of the total sludge disposal cost for Cane Run 4, 5 and 6.

^b Additive supply system cost (\$700,000) is shown separately. The \$700,000 is the 3/7 portion of the total additive supply system for SO_x removal systems for Cane Run 4, 5 and 6.

^c Includes plant overhead.

^d Erection equipment cost is included in plant overhead.

^e The capital investment for the above system in 1976 dollars is equivalent to \$15.95 million or 53.15 \$/kW (based on gross peak capacity).

Table 3-6
ESTIMATED AVERAGE ANNUAL OPERATING COSTS
FOR THE DUAL ALKALI PDS SYSTEM AT LOOSE CO.
(1979-1982)

Direct Costs:	Quantity	Unit Cost, \$	Annual Cost \$/yr
Materials			
Carbide Lime ^a	58,728 tons	13.29/ton	780,500
Soda Ash	1,912 tons	100.00/ton	191,200
Fuel Oil ^b	1,802,808 gals.	0.43/gal	775,200
Electricity	16,188,480 kwhr	0.01/kwhr	161,900
Water	126,100,000 gals.	0.05/1,000 gals.	6,300
Sludge Removal	185,280 tons	2.01/wet ton	372,400
Maintenance Materials			279,000
Labor			
Operation	26,280 hrs	8.18/hr	215,000
Maintenance	26,280 hrs	8.28/hr	217,600
Analysis	2,080 hrs	10.00/hr	20,800
Supervision	2,080 hrs	Various	40,000
Total Direct Costs			3,059,900
		No Reheat ^b	2,384,700
Indirect Costs:			
Overhead			293,100
Interest			1,064,500
Depreciation			724,700
Total Indirect Costs			2,082,300
Total Annual Operating Cost, \$			
H111s/kwhr		With Reheat	^b No Reheat
¢/10 ⁶ Btu		5,142,600	4,367,400
\$/Ton of S Removed		3.27	2.77
		32.8	27.9
		219.7	186.4

59.4% of 493,368 (Total Labor)
6.125% of 17,379,000
4.17% of 17,379,000

2.5% of total materials cost
from Table 3-4.

Basis: (a) If quick lime at \$40/ton is used as would be the case in most installations, the annual operating cost would increase from about \$4.4 to \$5.8 million. The corresponding increase in unit operating cost will be from 2.77 to 3.65 mills/kwhr.
(b) Operating cost for the no-reheat option was calculated without the cost of fuel oil.

Table 3-6
ESTIMATED AVERAGE ANNUAL OPERATING COSTS
FOR THE DUAL ALKALI FGD SYSTEM AT SIGECO
(1979 \$)(22)

	<u>Unit Cost, \$</u>	<u>Annual Cost, \$/Year</u>
Reagent Costs		
Pebble Lime	\$40/ton	1,210,000
Soda Ash	\$100/ton	162,000
Water	\$0.50/1,000 gals	41,000
Power Costs		
Demand Charge	\$700/kW	210,000
Energy Charge	\$0.015/kWhr	158,000
Capital Charge (@15% of Capital Cost)		1,500,000
Maintenance		180,000
Operating Labor (2/shift) ^(a)		240,000
Disposal Charge	\$2.00/ton	232,000
TOTAL		3,933,000/year 3.0 Mills/kWhr

Basis: 265 MW (Gross peak capacity) existing coal-fired plant, 60% load factor, 4 wt.% S in coal, 80 to 95% SO₂ removal, on-site disposal without treatment.

- (a) For actual operation 3 operators per shift will be required.
- (b) Total capital expended on unit estimated to be \$12,500,000 or \$50/kW.

Section 4

SUMMARY OF STATUS OF DUAL ALKALI TECHNOLOGY IN JAPAN⁽¹⁾

4.1 GENERAL BACKGROUND

About 42 major dual alkali FGD plants, having a combined capacity of about 4,000 MW equivalent, were operational in Japan at the beginning of 1978. Table 4-1 summarizes these systems by process developer, constructor, process type, and capacity. Approximately 45% of the total capacity represents utility boiler application (primarily oil-fired) while the remainder includes industrial boilers, sintering plants, smelters, and sulfuric acid plants. A summary of those applications has previously been presented in Table 1-2.

An interesting facet of emerging dual alkali technology in Japan is processes having unlimited oxidation tolerance and utilizing limestone as a regenerant. This is generally accomplished by circulating absorbents other than sodium sulfite. Examples include Dowa's basic aluminum sulfate process, Kawasaki's magnesium-gypsum process, and Kureha's sodium acetate-gypsum process. Some of these processes do not use a clear solution as the absorbent. The current EPA funded program at Gulf Power's Scholz Station is also presently investigating the feasibility of limestone regeneration with sodium sulfite systems. A brief discussion of the Dowa, Kawasaki, and Kureha processes follows.

4.2 DOWA BASIC ALUMINUM SULFATE FGD PROCESS⁽³³⁾

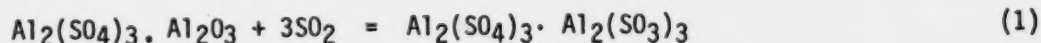
This process has been developed by Dowa Mining Co. of Japan. In the U.S. this process is licensed by Universal Oil Products (UOP) and is presently being tested at the TVA Shawnee Test Facility to evaluate its applicability to coal-fired boilers.

Process Description

As shown in Figure 4-1 this process consists of three operations: absorption, oxidation, and neutralization.

Absorption

SO₂ is absorbed in a clear solution of basic aluminum sulfate, Al₂(SO₄)₃ · Al₂O₃ of pH 3 to 4 to form Al₂(SO₄)₃ · Al₂(SO₃)₃;



Oxidation

The aluminum sulfite in the spent liquor is oxidized by air to aluminum sulfate:

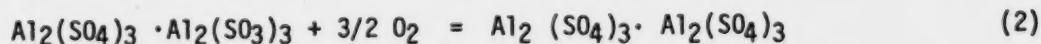
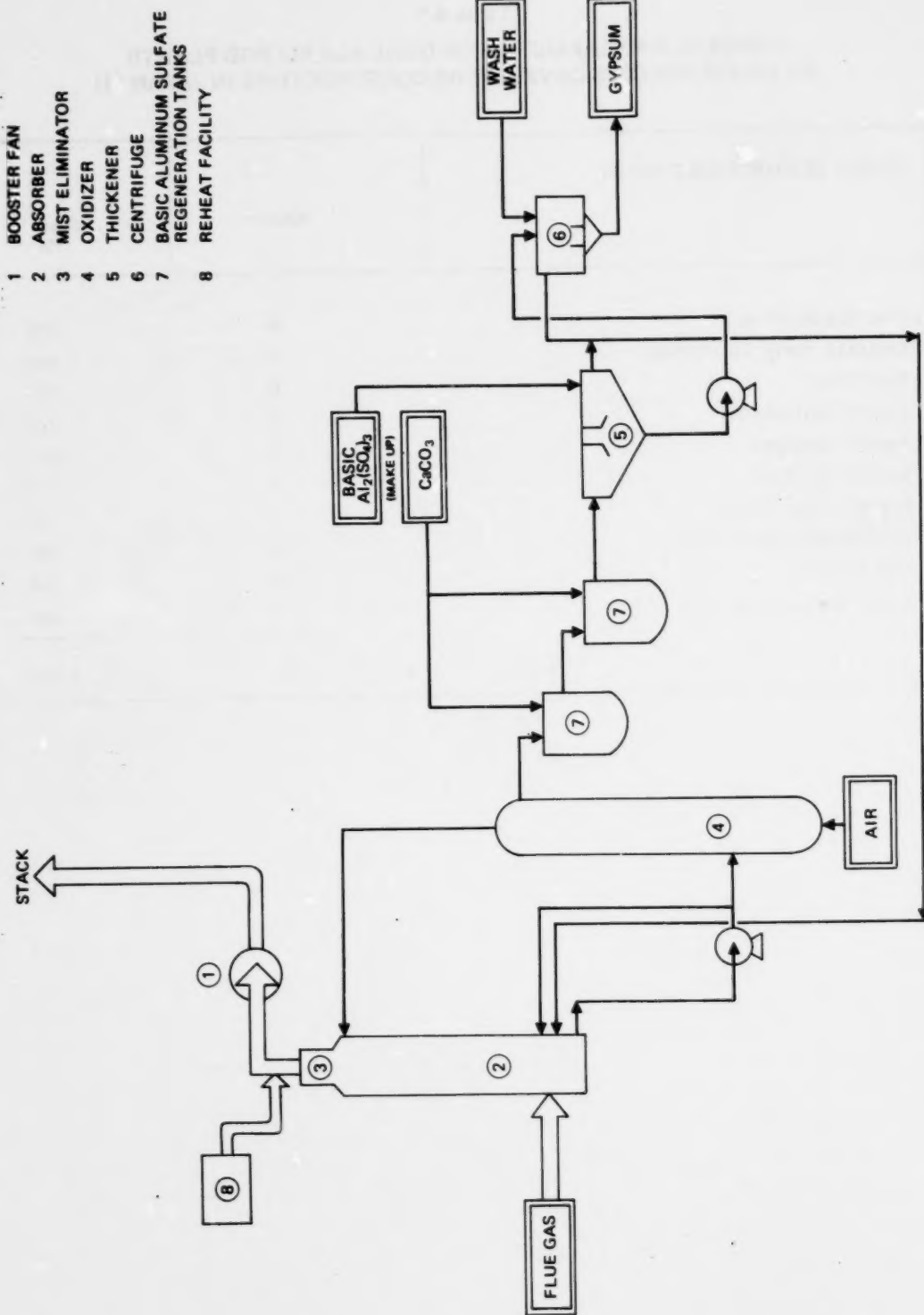


Table 4-1
NUMBERS AND CAPACITIES OF DUAL ALKALI FGD PLANTS
BY MAJOR PROCESS DEVELOPERS/CONSTRUCTORS IN JAPAN (1)

PROCESS DEVELOPER/CONSTRUCTOR	Number	Capacity MW
Dowa Engineering	8	200
Kawasaki Heavy Industries	2	180
Kobe Steel	6	745
Kurabo Engineering	5	215
Kureha Kawasaki	5	1685
Kureha Chemical	1	2
Nippon Kokan (NKK)	1	45
Tsukishima Kikai (TSK)	2	160
Showa Denko	3	360
Showa Denko-Ebara	9	370
TOTAL	42	3,962

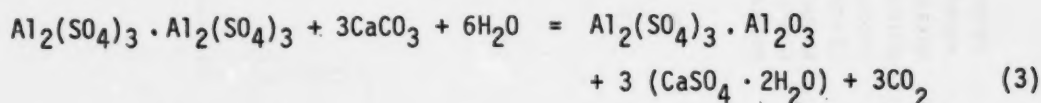


- 1 BOOSTER FAN
- 2 ABSORBER
- 3 MIST ELIMINATOR
- 4 THICKENER
- 5 CENTRIFUGE
- 6 BASIC ALUMINUM SULFATE REGENERATION TANKS
- 7 REHEAT FACILITY

Figure 4-1 DOWA BASIC ALUMINUM SULFATE - GYPSUM FLUE GAS DESULFURIZATION PROCESS

Neutralization

The oxidized liquor is treated with powdered limestone to precipitate gypsum and to regenerate the basic aluminum sulfate solution:



Commercial Applications

One of the earlier applications (~1 MW) was at the Mobara Works of Taenaka Mining. This plant started operation in October 1972 to treat waste gas from a molybdenum sulfide roaster containing 7500 ppm SO₂ at 100°C. No problems have been reported to date. Two commercial units, each with a capacity of treating 150,000 Nm³/hr (50 MW) of tail gas from a sulfuric acid plant were built at Okayama Works of Dowa. For these systems, inlet SO₂ averages 600 ppm. They are designed to remove at least 95% SO₂ at a liquid-to-gas ratio of 15-20 gal./10³ acf. These plants have operated successfully since startup in July 1974, with an average SO₂ removal efficiency of 99% during 20 months of testing. The only installation on a boiler, a small one at Naikai, also continues to perform well.

Other recent applications include two on iron ore sintering plants, one on a sulfuric acid plant, and one on a smelter.

Evaluation

The attractive features of this process, relative to the concentrated sodium-based processes, are unlimited oxidation tolerance and the use of limestone as a regenerant. However, this process operates at a higher liquid-to-gas ratio than the sodium-based processes.

4.3 KAWASAKI MAGNESIUM - GYPSUM FGD PROCESS(30)

Kawasaki Heavy Industries markets two versions of the magnesium-gypsum flue gas desulfurization process. In their "standard" process only a portion of the absorber bleed stream is oxidized and the mole ratio of calcium to magnesium in the absorbent liquor is maintained between 3 and 4. In the "new" process all the bleed stream is oxidized and the Ca/Mg mole ratio is held below 1.0.

Even though both the known commercial applications use the "standard" process, the description presented in this paper relates to the new process because of its greater potential. Figure 4-2 is a simplified process flow diagram of this "new" process.

Process Description

Unlike the Dowa process, this process does not employ clear liquor scrubbing.

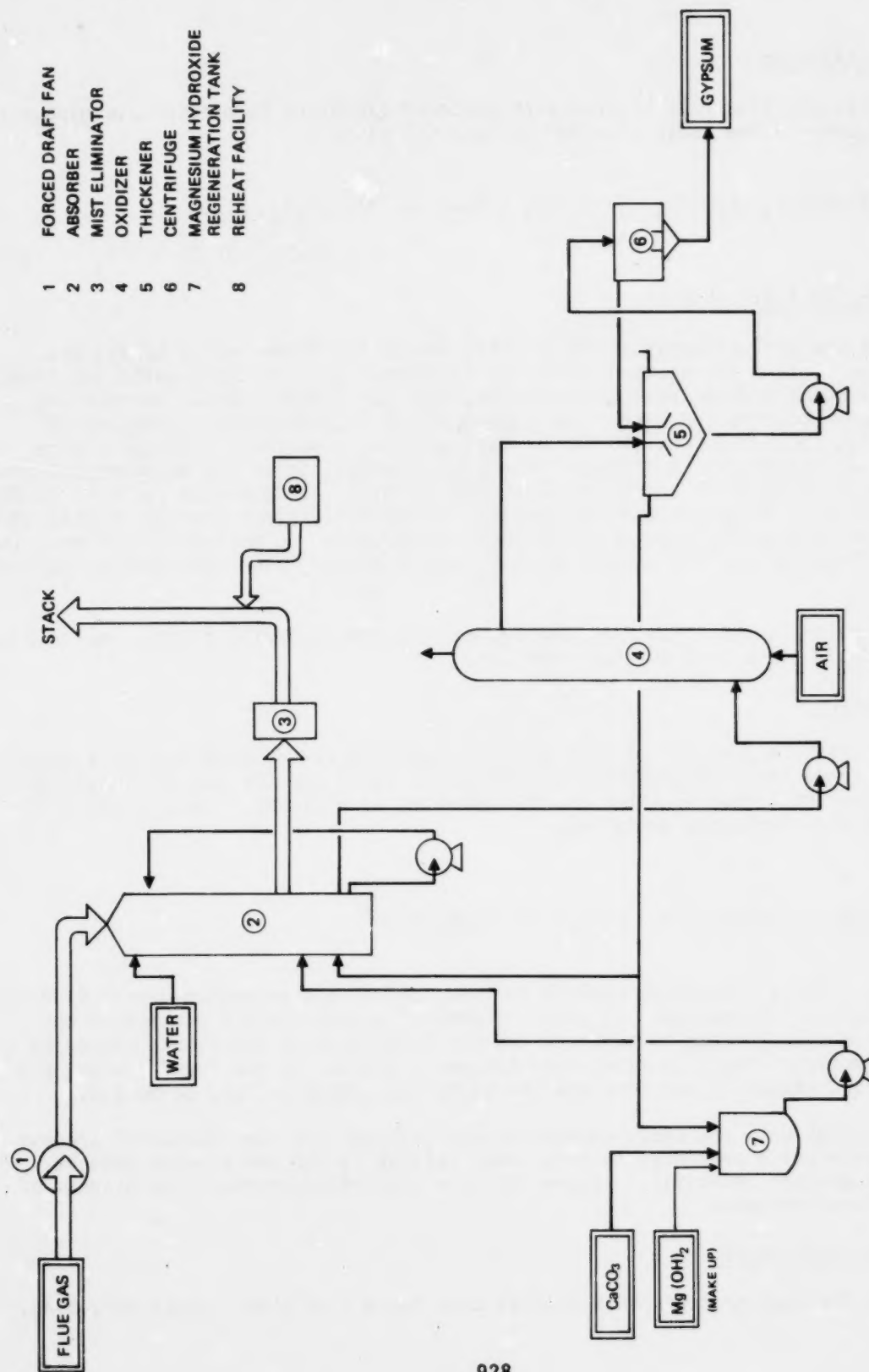
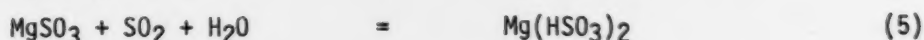
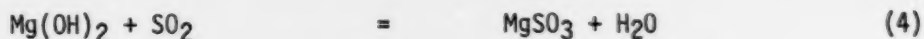


Figure 4-2 KAWASAKI MAGNESIUM - GYPSUM FLUE GAS DESULFURIZATION PROCESS (NEW)

Its three unit operations (absorption, oxidation, and gypsum recovery, and magnesium hydroxide regeneration) are described below.

Absorption

Flue gas containing SO_2 is contacted with a mixed slurry of calcium (as CaSO_4) and magnesium compounds. SO_2 is absorbed and removed according to reactions (4) and (5).



Oxidation and Gypsum Recovery

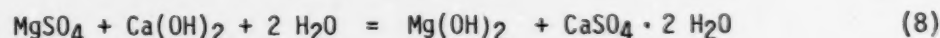
The spent liquor from the absorption section is oxidized by air to convert the magnesium sulfite to sulfate by the following reactions:



Since magnesium sulfate has a high solubility in water relative to calcium sulfate (calcium sulfate is recycled for seeding), separation of the calcium and magnesium sulfates is easy. Gypsum slurry from the oxidizer is dewatered to less than 10 wt.% moisture with a centrifuge. Magnesium sulfate solution is forwarded to the magnesium hydroxide regeneration section.

Magnesium Hydroxide Regeneration

Since magnesium sulfate has no capacity to absorb SO_2 , it is converted back to magnesium hydroxide. The regeneration is accomplished by adding lime or limestone. The following reaction occurs with lime.



Magnesium hydroxide thus regenerated is returned to the absorber along with the precipitated gypsum.

Commercial Applications

Kawasaki Heavy Industries has built two commercial plants which have been operational since the beginning of 1976. Both plants use the partial oxidation mode (standard process). The first commercial plant utilizing this process is the Okazaki Works of the Unitika Co.; the second is the Saidaji Works of

Japan Exlan Co. For the latter plant, inlet SO₂ averages 1400 ppm. For 95% SO₂ removal, a liquid-to-gas ratio of 35-45 gal./10³acf is required. The mole ratio of calcium to magnesium is maintained between 3.5 and 4.

Lime is used in the plant for Japan Exlan Co., whereas limestone is used as the main absorbent (with a little lime) for the Unitika Co. For the Unitika application, lime is added to the reactor, and limestone is added to the absorber; the ratio of limestone to lime is about 5 to 1.⁽¹⁾ The use of limestone is a Unitika Co. refinement. Gypsum is recovered as a by-product in both plants.

Evaluation

Like the Dow process, the major attractions of this process are unlimited oxidation tolerance and the option of limestone as a regenerant. However, the reported liquid-to-gas ratio for the "standard" version is significantly higher than that of the sodium-based processes. However, for the "new" version the liquid-to-gas ratio is comparable to that of the sodium-based process because most of the scrubbing is done by magnesium.

4.4 KUREHA SODIUM ACETATE-GYPSUM FGD PROCESS⁽²⁵⁾

Two limitations of the sodium sulfite dual alkali process marketed by Kureha were its inability to accommodate high oxidation levels and its difficulty in using limestone as a regenerant. In response to these shortcomings, Kureha developed the sodium acetate process described below.

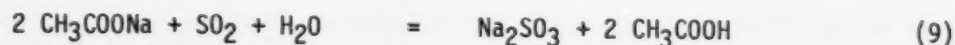
Process Description

The process is composed of three unit operations: absorption, oxidation, and gypsum recovery.

Absorption

As illustrated in Figure 4-3, the absorption tower consists of two sections: SO₂ absorption and an acetic acid recovery section linked in series in the same absorption tower.

In the SO₂ absorption section, SO₂ is removed by circulating sodium acetate solution. The following reaction takes place:



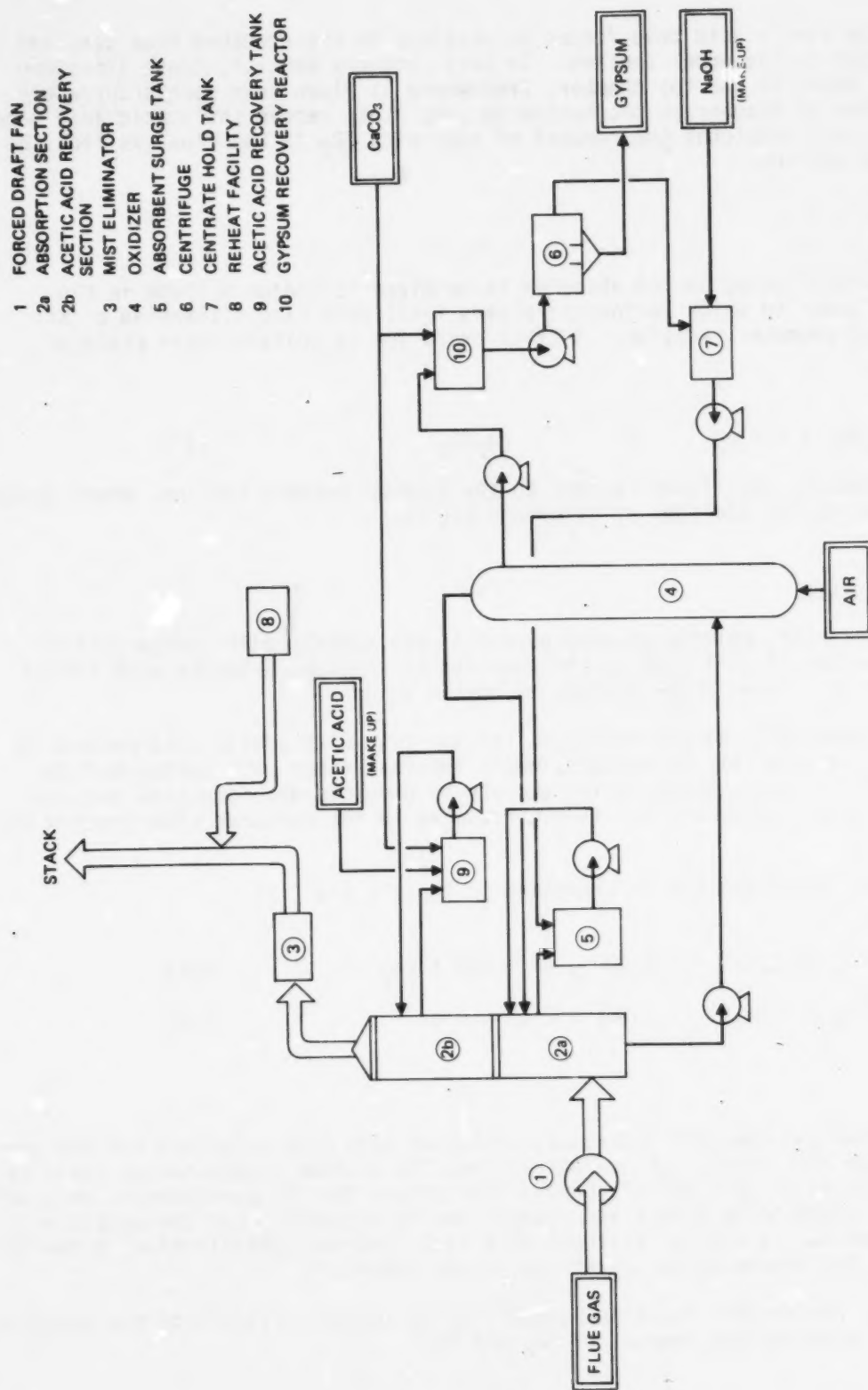
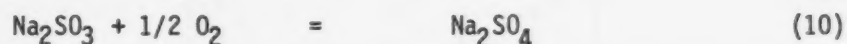


Figure 4-3 KUREHA SODIUM ACETATE - GYPSUM FLUE GAS DESULFURIZATION PROCESS

Part of the acetic acid thus formed volatilizes in the scrubbed flue gas, and is recovered in the upper section. In this recovery section, fresh limestone slurry is added to the top chamber, from where it flows down countercurrently from chamber to chamber to the bottom to completely remove the acetic acid vapor. This step also completes the removal of remaining SO₂ in the flue gas from the absorption section.

Oxidation

Sodium sulfite formed in the absorber is oxidized to sodium sulfate in the oxidation tower in which perforated plates facilitate fine dispersion of air bubbles and promote oxidation. Sulfite oxidation to sulfate takes place as follows:



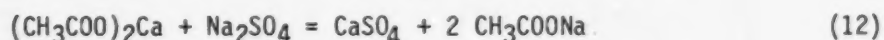
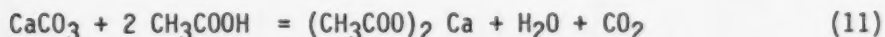
After oxidation, the liquor is sent to the gypsum recovery section, where gypsum is produced by the addition of limestone slurry.

Gypsum Recovery

In this operation, calcium carbonate reacts very rapidly with sodium sulfate in the presence of acetic acid; the reaction of calcium carbonate with sodium sulfate is very slow in the absence of acetic acid.

It is believed that calcium carbonate reacts first with acetic acid present in the liquor to form calcium acetate, which reacts further with sodium sulfate to form calcium sulfate and sodium acetate by a double-decomposition reaction. The regenerated sodium acetate is recirculated to the scrubber after separation of gypsum.

The reaction mechanism can be expressed as in (11) and (12):



Evaluation

Like the Dowa and Kawasaki processes, unlimited oxidation tolerance and the use of limestone as a regenerant are attractive. The claimed liquid-to-gas ratio is on the order of 10 gal./10³acf. Kureha also offers the "lime-regenerant version" which is claimed to be simple and cheaper due to elimination of the acetic acid recovery section, a smaller effluent hold tank, and the consolidation of the oxidation and the regeneration operations in one vessel.

Reportedly, the process can also be modified by adding catalysts to the scrubbing liquor for simultaneous removal of SO₂ and NO_x.

Commercial Applications

Unlike the Dowa and Kawasaki processes, there are no commercial applications of this process. However, this process has been tested in a bench-scale plant (100 Nm³/hr) with flue gas from an oil-fired boiler for 2 years since 1973. Furthermore, on the basis of the results in the bench-scale plant, a 5,000 Nm³/hr pilot plant was constructed in 1975. This pilot plant has operated smoothly and data necessary for scale-up have been obtained.

Section 5

THE EPA DUAL ALKALI RESEARCH, DEVELOPMENT, AND DEMONSTRATION PROGRAM

5.1 BACKGROUND

The EPA has been actively involved in the development of dual alkali technology since the Second International Lime/Limestone Wet-Scrubbing Symposium held in New Orleans in November 1971. Some of the incentive to develop this technology stemmed from a paper presented by R.J. Phillips⁽²⁰⁾ of General Motors (GM) concerning GM's laboratory and pilot plant work with a dilute mode dual alkali system. The results of the GM effort appeared very encouraging at that time in light of the difficulties being experienced with lime/limestone systems. The development of dual alkali technology by the EPA has followed an orderly progression of scale from laboratory to pilot plant to prototype to a full-scale utility demonstration of the process. In addition, EPA also funded a program to evaluate the full-scale dilute mode dual alkali system in operation at a GM industrial boiler system (32 MW equivalent).

Some initial laboratory work on regeneration chemistry was done in the EPA/IERL-RTP laboratories at Research Triangle Park, in addition to an initial feasibility study⁽²³⁾ which indicated that dual alkali systems might be somewhat lower in capital and operating costs than lime or limestone systems under certain circumstances.

After these initial studies, the EPA contracted with ADL to conduct a study of the dual alkali process. The scope of work included in the initial laboratory and pilot plant program was subsequently expanded to include prototype testing at the Scholz Plant of Gulf Power Company where a 20 MW FGD prototype system was constructed by CEA/ADL for Southern Services. The Southern Company (parent of Southern Services) provided the funds for the project.

Goals of the initial EPA dual alkali program were to:

- Demonstrate reliable system operation.
- Demonstrate high SO₂ removal, 95% desirable, with high-sulfur coal.
- Demonstrate environmentally acceptable sulfate removal schemes.
- Minimize soluble materials in disposable waste.
- Minimize moisture in disposable waste.
- Demonstrate closed-loop operation.
- Minimize costs.
- Minimize Ca⁺⁺ concentration in the scrubber.

These goals have been successfully achieved up to the prototype level.

5.2 LABORATORY PROGRAM

Areas of investigation in the laboratory program include:

- Regeneration of simulated scrubber effluents with lime and limestone.
- Sulfate removal by precipitation of a mixed crystal containing CaSO_3 and CaSO_4 with water of hydration.
- Sulfate removal by reaction of Na_2SO_4 , CaSO_3 , and sulfuric acid.
- Feasible ranges of sulfate, chloride, magnesium, and iron in solution.
- Settling characteristics of the product solids.
- Fixation of dual alkali product solids.
- Density, compactability, leachability, and permeability of fixed and unfixed solids.

5.3 PILOT PLANT PROGRAM

In the pilot plant both short- and long-term runs (5 weeks/run) have been conducted to examine various modes of dual alkali operation. These modes of operation include:

- Concentrated alkali, sulfuric acid treatment, lime.
- Concentrated alkali, two-stage reactor system, lime.
- Concentrated alkali, single reactor, lime.
- Dilute alkali with sulfite oxidation, lime.
- Concentrated alkali, multi-stage reactor, limestone regeneration.
- Extra high concentrated alkali, two-stage reactor system, lime.

LaMantia, et al.⁽¹⁵⁾ summarizes the above noted laboratory and pilot plant work.

5.4 PROTOTYPE PROGRAM

The results of 17 months of testing at the 20 MW prototype facility of Gulf Power are contained in the report entitled "Final Report: Dual Alkali Test and Evaluation Program, Vol. III. Prototype Test Program - Plant Scholz," by LaMantia, et al.⁽¹⁶⁾ Current activity at this facility involves preparation for testing

of limestone as a regenerant.

5.5 GM INDUSTRIAL BOILER FGD SYSTEM EVALUATION

The GM test program at the 32 MW equivalent system at the Chevrolet Transmission plant in Parma, Ohio, was conducted under an agreement between the EPA and GM. The test program design and some chemical analyses were performed for GM and the EPA, under contract, by ADL. The FGD system consists of four, Koch tray, stainless steel scrubbers, 32 MW equivalent, and a 40 MW equivalent regeneration system consisting basically of two tanks and two reactor-clarifiers. The boiler system consists of two 60,000 lb/hr (of steam) and two 100,000 lb/hr boilers. The GM system is a dilute mode dual alkali system.

The system was operated intermittently since startup in February 1974. Availability data for the system were not easily defined since there are four boilers and four scrubbers while the steam demand was frequently less than the capacity of one or two boilers. Thus, when there was a problem with one of the systems it was taken out of service and replaced by one of the stand-by units.

A good account of this system was presented by Dingo and Piasecki⁽⁷⁾ at the 1974 EPA FGD Symposium in Atlanta. In addition, the results of an EPA funded evaluation of this system by Arthur D. Little are also available.⁽¹¹⁾

5.6 FULL-SCALE UTILITY FGD SYSTEM DEMONSTRATION

As indicated earlier, Louisville Gas and Electric's Cane Run No. 6 unit was chosen as the site for a full-scale, EPA co-funded, dual alkali demonstration effort. The vendors providing the system and the monitoring program director are CEA/ADL and Bechtel National, Inc., respectively.

The program, to be conducted on this 277 MW unit, consists of four phases:

- (I) Process design and cost estimate.
- (II) Engineering design, construction, and mechanical testing.
- (III) Startup and acceptance testing.
- (IV) One year operation and long-term testing.

As of this date, phases I and II have been completed. Phase III is presently in progress. Results of phase I have been reported by Van Ness, et al.⁽³²⁾

Summary design criteria for this demonstration unit include:

- Unit will meet all applicable pollution control regulations when the pulverized coal-fired boiler burns 2.5 - 4.5 wt.% sulfur coal.
- Instrumentation will allow accurate material and energy balances.

- SO₂ outlet will be held below 200 ppm for coal sulfur less than 5 wt.%. For higher sulfur coals, SO₂ removal will be at least 95%.
- Stack gas will be reheated.
- Chemical makeup levels:
 - Na = 0.0495 mole Na or monovalant Cation/mole S
 - Ca = 1.01 mole Ca/mole S

An extensive monitoring program will be conducted at this facility. The program will characterize effluents, document chemical consumption, evaluate sludge disposal, and evaluate system reliability and process economics.

5.7 FUTURE PLANS

Based on the level of future funding, the EPA dual alkali program may be upgraded and supplemented by expanding the currently planned monitoring program on the Louisville Gas and Electric's 280 MW facility by 1 year to allow testing of: (1) limestone as an alternative to the more expensive, energy-intensive lime reactant; (2) methods for upgrading the quality of sludge produced and to compare disposal options; and (3) strategies to control the multimedia environmental impact of all effluents and emissions.

In addition, process evaluation programs may be initiated on the other two full-scale dual alkali facilities presently in the final stages of construction, if appropriate contractual arrangements can be made with the respective utility company and process supplier. These facilities are the 575 MW Central Illinois Public Service and 250 MW Southern Indiana Gas and Electric systems being engineered by Envirotech and FMC, respectively. Parallel test programs may be conducted by an independent EPA subcontractor to evaluate, characterize, and compare these full-scale facilities, and allow evaluation of promising dual alkali process variations.

Section 6

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Appendix A

DEFINITION AND DISCUSSION OF TERMS(13)

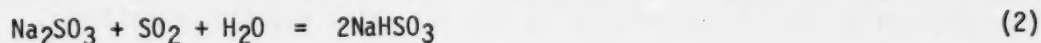
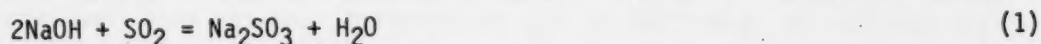
As with any specialized technology, a discussion of flue gas desulfurization in general, and dual alkali technology more specifically, involves the use of special terminology which has evolved with the technology. While terms are understandable to those dealing with the subject on a daily basis, they can be somewhat ambiguous to others. To clarify some of these ambiguities, and to define terms used here and by others describing dual alkali technology development, a number of terms and concepts are defined and discussed in a general sense.

A.1 ABSORPTION/REGENERATION CHEMISTRY

The main chemical reactions that take place in dual alkali systems can be divided functionally into the absorption and regeneration reactions. A number of secondary reactions which have very important effects on the overall functioning of the system also take place. These include oxidation, softening, and sulfate removal reactions which are discussed under the appropriate headings.

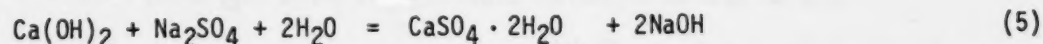
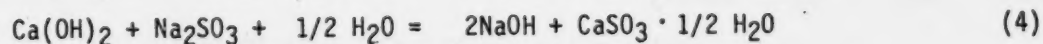
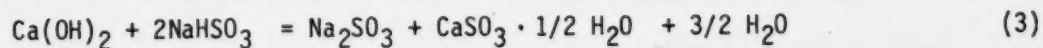
The regeneration reactions and in some cases the absorption reactions will depend upon which calcium supplying regenerant is used--lime or limestone. With lime, the system can be operated over a wider pH range than with limestone. This wider pH range allows lime systems to operate over the complete range of active alkali hydroxide/sulfite/bisulfite, whereas limestone systems can only operate in the sulfite/bisulfite range.

The main overall absorption reactions are described by the following equations:

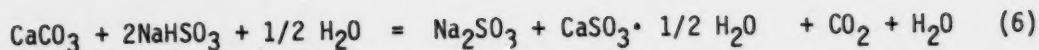


The main overall regeneration reactions are described by the following equation for lime and limestone, respectively:

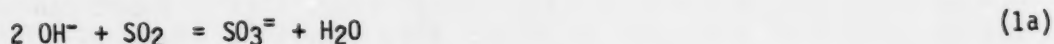
Lime



Limestone

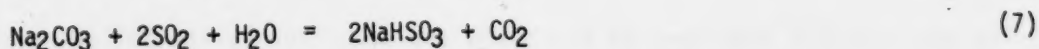


D/A systems are frequently and erroneously referred to as sodium ion scrubbing systems. It should be stated that, from a "pure" chemistry viewpoint, the reactions presented in equations (1-11) and (15-18) do not involve the sodium ion (Na^+); however, the presentation is made using compounds of sodium because sodium systems are prevalent in D/A applications and because this allows showing the reactions using electrically neutral reactants and products rather than charged ions. For example, the absorption reactions involve reaction of SO_2 with an aqueous base such as OH^- , $\text{SO}_3^{=}$, HCO_3^- , or $\text{CO}_3^{=}$, rather than with Na^+ which does not take part in the reaction, but is only present to maintain electrical neutrality. Thus equations (1) and (2) for example could have been written as equations (1a) and (2a), respectively:

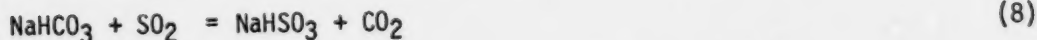


A.2 ACTIVE ALKALI

This term refers to the concentrations of NaOH , Na_2CO_3 , NaHCO_3 , Na_2SO_3 , and NaHSO_3 in the scrubbing solutions. Sodium bisulfite is included in this definition although it is not technically an alkali (i.e., it cannot react with SO_2 in these systems); however, it can be converted to an alkali by reaction with lime or limestone. It should also be noted that the molar capacity of each of these species for absorption of SO_2 is different, and can vary from zero to 2 moles of SO_2 per mole of active alkali. This difference in molar capacity for absorption of SO_2 is illustrated by the following reaction equations:



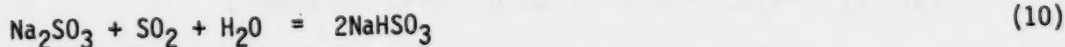
(sodium carbonate molar capacity: 2 moles SO_2 /mole)



(sodium bicarbonate molar capacity: 1 mole SO_2 /mole)



(sodium hydroxide molar capacity: 1 mole SO_2 /mole)



(sodium sulfite molar capacity: 1 mole SO_2 /mole)



(11)

(sodium bisulfite molar capacity: zero mole SO_2 /mole)

Molar capacity is simply the number of moles of SO_2 needed to convert 1 mole of the absorbent alkali completely to sodium bisulfite. Since there is a difference in the molar capacity of different alkali components to absorb SO_2 , active alkali is a descriptive rather than a quantitative term. If the concentration of each active alkali component (moles/liter) is known, the capacity of the scrubbing liquor to absorb SO_2 (moles of SO_2 /liter of solution) can be calculated as the sum of each of the active alkali component concentrations multiplied by their respective molar capacities as follows:

Scrubber liquor SO_2 capacity (moles/liter) =

$$2 [\text{Na}_2\text{CO}_3] + [\text{NaOH}] + [\text{NaHCO}_3] + [\text{Na}_2\text{SO}_3]$$

A.3 TOS

Total oxidizable sulfur (TOS) denotes the concentration of sulfur compounds in solution in which the sulfur is in the +4 oxidation state. Simply, this is the total concentration of sulfite plus bisulfite.

$$\text{TOS (moles/liter)} = [\text{SO}_3^{2-}] + [\text{HSO}_3^-]$$

Sulfate is not part of TOS, since the sulfur is in the +6 oxidation state in this species. Sulfur dioxide dissolving in scrubbing solutions increases the TOS in solution.

A.4 ACTIVE SODIUM

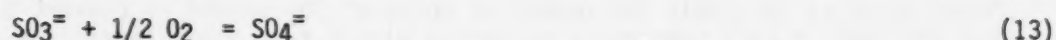
This is the concentration of sodium in solution which is associated with the active alkali.

$$[\text{Na}^+]_{\text{act}} = [\text{NaOH}] + 2 [\text{Na}_2\text{CO}_3] + [\text{NaHCO}_3] + [\text{NaHSO}_3] + 2 [\text{Na}_2\text{SO}_3]$$

If NaOH , NaHCO_3 , Na_2CO_3 , NaHSO_3 , or Na_2SO_3 solids are added to dual alkali solutions, the increase in sodium ion in solution is "active sodium." If Na_2SO_4 or NaCl , for example, is added, the increase in sodium ion is "inactive sodium." Active sodium is not increased by the dissolution of SO_2 in scrubber solutions. Note that the term "active sodium" can be misleading in that the sodium ion doesn't participate in any of the process reactions.

A.5 OXIDATION

Oxidation in a dual alkali system refers to the conversion of TOS to sulfate by one of the following equations:



Simple oxidation of SO_2 to SO_3 in the flue gas is also considered oxidation in the dual alkali system:



Oxidation in the system has the effect of changing active sodium to inactive sodium, or active alkali to inactive alkali.

Oxidation may occur in any part of the system: in the scrubber, the reaction vessels, or in the solids separation equipment. In general, the rate of oxidation in the system is thought to be a function of the rate of dissolution of oxygen, pH of the scrubbing solution, impurities present in solution, and concentration of reactants. Oxidation rate is thus affected by composition of the scrubbing liquor (scrubbing liquors containing high concentrations of dissolved salts may absorb oxygen more slowly), oxygen content of the flue gas, impurities in the coal and lime or limestone, and the design of the equipment (the regeneration and solids separation sections of the system in particular can be designed to limit dissolution of oxygen and the number of scrubber contact stages is extremely important).

Oxidation rate is expressed as a percentage and is calculated from an overall material balance on the system:

$$\text{Oxidation rate (\%)} = \frac{[\text{SO}_4^{=2-} \text{ leaving the system (moles)}]}{[\text{Total sulfur collected (moles)}]} \times 100$$

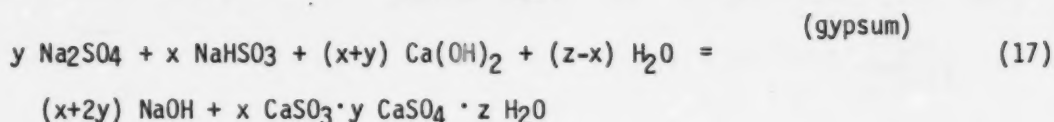
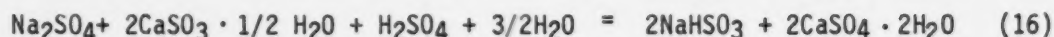
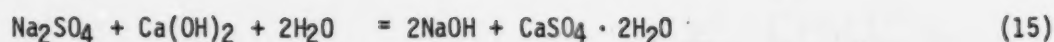
Sulfate leaving the system is total moles of sulfate in the solid waste plus any sulfate in the associated liquor.

A.6 SULFATE REGENERATION

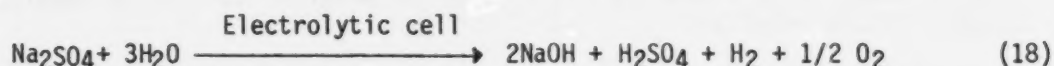
This term is a misnomer. What is really meant is sulfate removal from the system with regeneration of active alkali from inactive sodium sulfate. (See Sulfate Removal, below.)

A.7 SULFATE REMOVAL

Sulfate is removed from the system with regeneration of active alkali from inactive sodium sulfate. Examples of these sulfate removal reactions are:



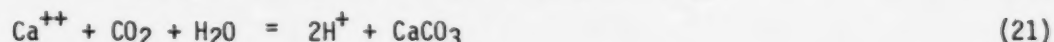
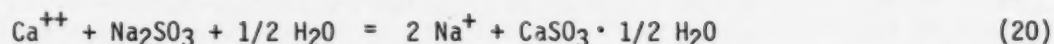
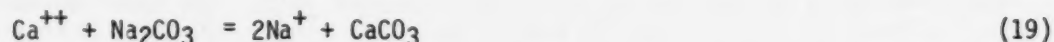
(mixed crystal or solid solution)



Sulfate should be removed in an environmentally acceptable manner; a simple purge of soluble Na_2SO_4 from the system to land or waterway disposal is not acceptable.

A.8 SOFTENING

This term is used to describe various methods used to lower the dissolved calcium ion concentration in regenerated solutions. The purpose of softening the scrubbing liquor before recycling to the scrubber is to ensure that it is subsaturated with respect to gypsum. This reduces the gypsum scaling potential in the scrubber. Examples of softening reactions are:



In each of the above reactions, calcium ions are removed from solution as part of an insoluble material, outside the scrubber system. Reactions (19) and (21) are referred to as carbonate softening. Reaction (20) is considered sulfite softening. Generally, dilute systems employ carbonate softening; concentrated system inhibit scaling due to the high sulfite concentrations which prevent high Ca^{++} ion concentrations in the scrubber liquor.

A.9 DILUTE VS. CONCENTRATED SYSTEMS

Dilute or concentrated refers to the active alkali concentration in a particular system. This differentiation is made because, in theory at least, based on their solubility products in water, both CaSO_3 and CaSO_4 should not precipitate from a solution of sulfite and sulfate simultaneously when using relatively small quantities of lime slurry for regeneration, unless the concentrations of sulfite and sulfate are present in a certain ratio. This can be shown by dividing one solubility product equation by the other:

$$\frac{[Ca^{++}] [SO_4^{--}]}{[Ca^{++}] [SO_3^{--}]} = \frac{K_{sp}}{K_{sp}}$$

from this, cancelling Ca^{++} ion concentration,

$$\frac{[SO_4^{--}]}{[SO_3^{--}]} = \text{constant}$$

The ratio of sulfate to sulfite for simultaneous precipitation of $CaSO_4$ and $CaSO_3$ is shown to be a constant. It is on the order of 10^3 - 10^4 in ideal, extremely dilute solutions.

The constant in the above equation is the ratio of solubility product constants of calcium sulfate and sulfite. Thus, in theory, if the ratio of sulfate to sulfite is higher than this constant, only calcium sulfate should precipitate; and if the ratio is lower than the constant, only calcium sulfite should precipitate.

This very simplified consideration of the chemistry given above is clouded in the "real world" by factors that contribute to non-ideal behavior of these systems. These factors include changes in ionic activities in solutions containing high electrolyte concentrations, and evidence of coprecipitation of calcium sulfite and sulfate in the form of a "mixed crystal" or "solid solution" in a manner which is not completely understood at present.⁽⁹⁾⁽¹⁴⁾

With due consideration to the non-ideal behavior of these systems, however, under given conditions, a ratio of sulfate to sulfite in solution can be determined at which the previously cited examples hold. The ratio establishes a definition for "dilute" or "concentrated" dual alkali systems. When the ratio is such that gypsum alone or both gypsum and calcium sulfite will precipitate from the solution with the addition of slaked lime, the system is "dilute."

A.10 LIME OR LIMESTONE STOICHIOMETRY

Lime or limestone stoichiometry can be expressed as a percentage based on an overall material balance around the system:

$$\text{Lime Stoichiometry} = \frac{\text{moles CaO added}}{\text{mole sulfur collected}} \times 100$$

$$\text{Limestone Stoichiometry} = \frac{\text{moles CaCO}_3 \text{ added}}{\text{mole sulfur collected}} \times 100$$

Lime or limestone stoichiometry is an indication of the efficiency of utilization of lime or limestone used. Ignoring alkali components in the fly ash collected and the alkalinity added with sodium makeup to the system, 100% stoichiometry is complete utilization; stoichiometries over 100% represent less

efficient utilization of lime or limestone. Stoichiometries under 100% indicate alkalinity from other sources.

A.11 FEED STOICHIOMETRY

This is calculated by a material balance around the scrubber. It is usually expressed as the ratio:

$$\text{Feed Stoich} = \frac{\text{Liquor SO}_2 \text{ Capacity (moles/liter)} \times \text{Flow (liter/min)}}{\text{SO}_2 \text{ (mole/min)}}$$

This ratio is evaluated for the gas and liquid streams entering the scrubber.

Feed stoichiometry is a measure of the ability of the incoming liquor to react with or absorb all of the incoming SO₂ in the scrubber, assuming ideal contact of gas and liquor. Feed stoichiometry above 1.0 is required for high SO₂ removal capability. At feed stoichiometry at or below 1.0, assuming ideal contact between the gas and liquor, there will be significant equilibrium SO₂ partial pressure above the liquor, and thus SO₂ removal is theoretically limited to the value calculated on the basis of this SO₂ partial pressure in the exiting flue gas.

Appendix B

DESIGN, PROCESS AND COST CONSIDERATIONS⁽¹³⁾

A commercial dual alkali system must be designed to remove the desired quantity of sulfur oxides from a given flue gas stream, while operating in a reliable manner and discharging environmentally acceptable solid waste product. In fulfilling these design objectives, cost is also important.

B.1 SO₂ REMOVAL

For some time it has been known that small quantities of sulfur dioxide can be removed from large amounts of relatively inert gas by cyclic processes involving absorption into aqueous solutions of sodium sulfite/bisulfite. Johnstone et al. published a paper in 1938 giving data on the vapor pressure of SO₂ over solutions of sulfite/bisulfite and methods of calculating these equilibrium values under various conditions. The equilibrium partial pressure of SO₂ over sulfite/bisulfite solutions, the theoretical limit which a practical design can approach, is generally a function of solution temperature, pH, concentration of sulfite/bisulfite, and total ionic strength. Since Johnstone's work, a number of organizations have pursued this technology with laboratory, pilot plant, and full-scale applications for flue gas desulfurization, and many have demonstrated its ability for high removal efficiencies. (Note that, although Johnstone's work was aimed at cyclic processes with thermal regeneration, such as the Wellman-Lord system, the vapor pressure data are also applicable to dual alkali systems which use chemical regeneration.)

Once methods have been established to determine equilibrium SO₂ vapor pressure over scrubbing solutions of the various concentrations to be encountered in an operating system, it becomes a matter of standard chemical engineering practice to design adequate gas absorption equipment to accomplish the desired SO₂ removal in a system. For comparison, note that the design of lime/limestone slurry absorption equipment is further complicated by the kinetics of dissolution of the lime or limestone, the particle size of the suspended material, and the crystal morphology of the lime or limestone.

B.2 RELIABLE OPERATION

System reliability can be adversely affected by two classes of problems: mechanical and chemical.

Mechanical problems include malfunction of instrumentation and mechanical and electrical equipment such as pumps, filters, centrifuges, and valves. These problems in a commercial FGD system can be minimized by careful selection of materials of construction and equipment and by providing spares for equipment items such as pumps and motors which are expected to be in continuous operation and are prone to failure after a relatively short period of operation. Another

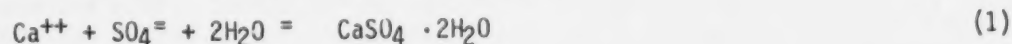
important consideration in minimizing mechanical problems is the institution of a good preventive maintenance program.

Chemical (or physical/chemical) problems which may be associated with a dual alkali system include scaling, production of poor-settling solid waste product, excessive sulfate buildup, water balance, and buildup of non-sulfur solubles which enter the system as impurities in the coal or lime. Each factor is associated with reliable system operation, or production of an environmentally acceptable solid waste.

Scaling - One of the primary reasons, and probably the most important, for development of dual alkali processes was to circumvent the scaling problems associated with lime/limestone wet scrubbing systems. Therefore, a dual alkali system should be designed to operate in a non-scaling manner.

Scaling is caused by precipitation of calcium compounds (from process liquors) on the surfaces of various system components. In the scrubber, this is particularly troublesome since the flue gas path through the scrubber, if affected, could shut down the boiler/scrubber system and lower reliability.

Since scrubbing in dual alkali systems employs a clear solution rather than a slurry, there is a tendency to ignore potential scaling problems. Testing experience with dual alkali systems has indicated, however, that scaling can occur and indeed the problem should be a legitimate concern in the design of any system. Both gypsum and carbonate scale buildup has been recognized in these systems. Gypsum scaling is caused by the reaction of soluble calcium ion with sulfate ion formed in the system through oxidation of the absorbed SO_2 or from absorbed SO_3 according to the reaction:



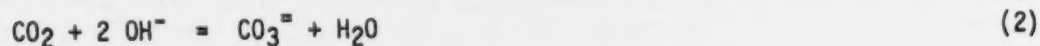
In dilute systems, gypsum scaling is controlled by softening the regenerated liquor prior to recycling to the scrubber. In concentrated systems, gypsum is not a problem since the high sulfite concentration keeps the Ca^{++} ion concentrations low. Softening ensures that the liquor recycled to the scrubber system is unsaturated with respect to gypsum; therefore, with proper softening even if some sulfate is formed in the scrubber, the liquor will not be saturated with gypsum and cause scaling on the inside surfaces of the scrubber. In concentrated active alkali systems, a special softening step is not necessary since high sulfite concentration is maintained throughout the system. This sulfite maintains a low Ca^{++} ion concentration (sulfite softening), and thus maintains the scrubbing solution unsaturated with respect to gypsum.

Based on experience gained in lime/limestone scrubbing testing, a certain factor of safety in the prevention of gypsum scaling probably exists in dual alkali systems. Gypsum has been found to supersaturate easily to about 130% saturation. Thus, scaling will occur only if supersaturation is excessive.

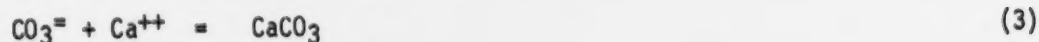
Carbonate scaling usually occurs as a result of localized high pH scrubbing liquor in the scrubber where CO_2 can be absorbed from the flue gas to produce CO_3

ions. These ions subsequently react with dissolved calcium to precipitate calcium carbonate scale according to the following series of reactions:

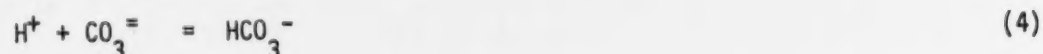
Carbon dioxide absorption by high pH liquor:



Calcium carbonate scaling:



Based on experience with the General Motors, full-scale, dual alkali system, carbonate scaling could occur with scrubber liquor pH above 9. At lower pH, the carbonate/bicarbonate equilibrium system tends to limit the free carbonate ion and thus prevent precipitation of calcium carbonate:



Thus, carbonate scaling can be eliminated by control of pH in the scrubber.

Solids Quality - Under certain conditions, the waste solids produced in the regeneration sections of various dual alkali systems have a tendency not to settle from the scrubber liquors. This creates problems in the operation of settlers, clarifiers, reactor clarifiers, filters, and centrifuges. Although observed in the laboratory testing conducted by the EPA on dilute systems and in the laboratory and pilot plant work conducted by ADL on dilute and concentrated systems, this phenomenon is not completely understood, but is thought to be a function of reactor kinetics.⁽¹⁴⁾

Some of the factors thus far identified which appear to affect the solids settling properties are reactor configuration, concentration of soluble sulfate, concentration of soluble magnesium and iron in the liquor, concentration of suspended solids in the reaction zones, and use of lime vs. limestone for reaction. Based on laboratory work in dilute systems (about 0.1 M active sodium) using limestone, it appears that solids settling characteristics degraded significantly at soluble sulfate levels above 0.5 M. Based on laboratory work with concentrated systems (about 0.45 M TOS, 5.4 pH, 0.6 M sulfate) using limestone, marked degradation of solids settling properties occurred at a magnesium level of 120 ppm and virtually no settling of solids occurred at the 2000 ppm magnesium level. Equal degradation of solids settling properties also occurred in concentrated systems when the sulfate level was raised to 1.0 M while maintaining low magnesium level (about 20 ppm) and keeping other variables constant.⁽¹⁴⁾

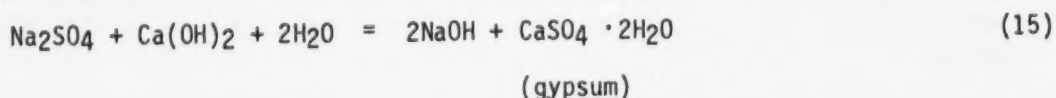
Envirotech⁽⁴⁾ advocates the recycle of precipitated solids from the thickener underflow to the reaction zones in an effort to grow crystals which settle faster and are more easily filtered.

ADL cites reactor configuration as being important in the production of solids with good settling and filtration characteristics.⁽¹⁴⁾ Their basis for this is comparative tests of a simple continuous flow stirred tank reactor (CFSTR) with the ADL/Combustion Equipment Associates (ADL/CEA) designed reactor system under similar conditions. The ADL/CEA reactor system appeared to give better settling solids over a greater range of conditions than a simple CFSTR. The ADL/CEA reactor system consists of a low residence unit followed by a high residence unit.

Sulfate Removal - In dual alkali systems, some of the sulfur removed from the flue gas takes the form of soluble sodium sulfate due to oxidation in the system, thus changing some of the active sodium to the inactive variety. When sodium in the system is converted to the inactive form (Na_2SO_4), it is relatively difficult to convert back to active sodium. To convert inactive sodium to active sodium, sulfate ion must be removed from the system in some manner, while leaving the sodium in solution. The alternative to this is to remove the sodium sulfate from the system at the rate it is being formed in the system. This alternative is not desirable since it wastes sodium and generally is carried out by allowing the sodium sulfate to be purged from the system in the liquor which is occluded in the wet solid waste product.⁽¹⁴⁾ The solid waste product can then potentially contribute to water pollution by leaching. Water runoff can contaminate surface water, while leaching and percolation of the leachate into the soil can contaminate ground water near the disposal site. Failure to allow for sulfate removal from dual alkali systems will ultimately result in a) precipitation of sodium sulfate somewhere in the system if active sodium is made up to the system, or b) in the absence of makeup, eventual deterioration of the SO_2 removal capability due to the loss of active sodium from the system.

Equations (15), (16), (17), and (18), shown previously under the definition of "sulfate removal" in Appendix A, describe several sulfate removal techniques which have been used in FGD system pilot tests.

The first equation depicts the sulfate removal technique used in dilute active alkali systems:



Concerning the full-scale dilute alkali system installed and operating at the Parma, Ohio, transmission plant of General Motors, and dilute systems in general, Phillips⁽¹⁹⁾ stated:

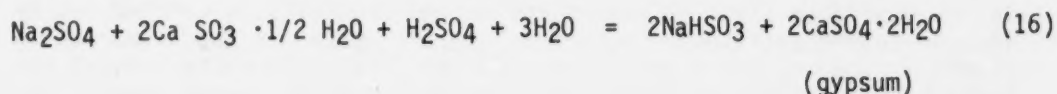
"The presence of Na_2SO_4 in the scrubber effluent is the prime factor influencing the design of the regeneration system. Na_2SO_4 is not easily regenerable into NaOH using lime, the reason being that the product, gypsum, is relatively soluble . . . Na_2SO_4 cannot be causticized in the presence of appreciable amounts of SO_3^{2-} or OH^- because Ca^{++} levels are held below the CaSO_4 solubility product. To provide for sulfate causticization, the system must be operated at dilute OH^- concentrations below

0.14 molar. At the same time, $\text{SO}_4^{=}$ levels must be maintained in the system at sufficient levels to effect gypsum precipitation . . . We selected 0.1 molar OH^- and 0.5 molar $\text{SO}_4^{=}$ as design criteria."

In a previous paper, Phillips(20) showed a plot of equilibrium caustic formation in $\text{Ca}(\text{OH})_2$ - Na_2SO_4 solutions at 120°F which is the basis for selection of the design criteria. The essence of this discussion is that, if the active sodium concentration is sufficiently dilute, sulfate can be removed from the system by simple precipitation as gypsum by reaction of lime with sodium sulfate.

Since, as explained above, this reaction will not proceed to a great extent in concentrated active alkali systems, other techniques must be employed to effect sulfate removal in these systems.

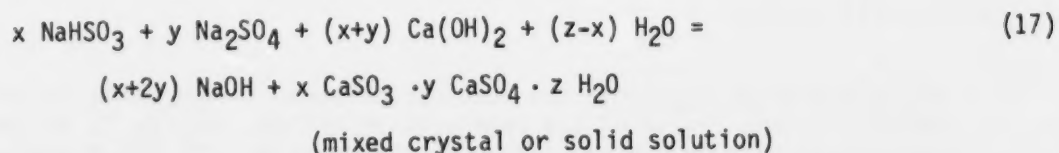
The second equation depicts a technique which is used in the full-scale dual alkali systems in Japan, and which has been pilot-tested by ADL under contract with EPA:



This technique is used to precipitate gypsum by dissolving calcium sulfite in acidic solution, thus increasing the Ca^{++} in solution enough to exceed the solubility product of gypsum. Ideally according to equation (16) 2 moles of gypsum should be precipitated for each mole of sulfuric acid added. In practice, however, this is not the case since any material which functions as a base can consume sulfuric acid and reduce the efficiency of this reaction for its intended purpose.(14) Unreacted lime or limestone, sulfite ion, and even sulfate ion can consume sulfuric acid, thus lowering sulfate removal from the system.

Conceivably, this method of sulfate removal may be economically unattractive in applications with very high oxidation rates, and where the gypsum produced must be discarded. The economic picture is considerably changed where this system is used merely as a slipstream treatment to supplement other sulfate removal methods and/or where the solid product gypsum is saleable as is the case in Japan.

The third equation describes a phenomenon which has been referred to as mixed crystal or solid solution formation:



This phenomenon is described by EPA/IERL-RTP's R.H. Borgwardt(9) as it applies to lime/limestone wet scrubbing based on pilot plant investigations. A similar phenomenon has been observed by ADL in some of their early pilot testing of dual alkali systems in conjunction with CEA, and later in the EPA/ADL dual alkali test program.

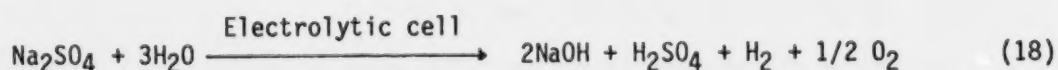
Under certain conditions the solids precipitated in lime/limestone and dual alkali systems contain sulfate, sulfite, and calcium; however, the liquor from which these solids precipitate appears to be subsaturated with respect to gypsum. This is based on the fact that pure gypsum crystal could be dissolved in the mother liquor from which the mixed crystal/solid solution was precipitated. In addition, the solid material, examined by X-ray diffraction contained no gypsum; infrared analysis confirmed the presence of sulfate.

Borgwardt found that the molar ratio of sulfate to sulfite in these solids was primarily a direct function of sulfate ion activity in the mother liquor. In pilot test work with lime/limestone scrubbing, with little or no chlorides present and normal magnesium level (below 1000 ppm) in solution, the sulfate to sulfite molar ratio in the mixed crystal/solid solution was precipitated. In addition, the solid material, examined by X-ray diffraction contained no gypsum; infrared analysis confirmed the presence of sulfate. This is equivalent to a $[SO_4^{2-}]/\text{total } [SO_x]$ ratio in the solids of 0.19.

In pilot test work with concentrated dual alkali systems, ADL observed the simultaneous precipitation of sulfate and sulfite with calcium in lime and limestone treatment of concentrated dual alkali scrubbing liquors. This phenomenon was surprising at first, in light of the reasoning which led to the development of dilute dual alkali systems; i.e., gypsum cannot be precipitated from solutions containing high active alkali concentrations. It was a simple technique for sulfate removal in concentrated systems. The $[SO_4^{2-}]/\text{total } [SO_x]$ ratio observed in pilot dual alkali work was as high as 0.20.⁽¹⁴⁾ Coincidentally, this was the same value observed by Borgwardt in lime/limestone testing. This led to the belief that the same phenomenon was occurring in both processes. The mother liquor from which these solids were precipitated was also found to be subsaturated in gypsum, and when the solids were examined, pure gypsum was not found.

Based on the observed data, it appears reasonable to design a concentrated active alkali system for a particular situation in which the system oxidation rate is below about 20%.⁽¹⁴⁾ In this case, sulfate can be removed at the desired rate, without purging Na_2SO_4 or supplementing the system with other complex methods of sulfate removal.

The fourth equation shows sulfate removal as sulfuric acid in an electrolytic cell:



This method is the basis for operation of the Stone and Webster/Ionics process sulfate removal technique. In Japan, Kureha/Kawasaki has pilot-tested the Yuasa/Ionics electrolytic process for sulfate removal in conjunction with their dual alkali process. They claim that this process is less expensive overall than the presently used sulfuric acid addition method. In addition, they claim that sodium losses from the system can be cut in half through the use of this method: from 0.018 moles Na loss/mole SO_2 absorbed to 0.009 moles Na loss/mole SO_2 absorbed.⁽²⁴⁾

Another approach to sulfate control is to limit oxidation. With sufficient limitation of oxidation, by process and equipment design, it may be possible to control sulfate by a small unavoidable purge of Na_2SO_4 with the solid waste product. To design for minimum oxidation, there should be minimum residence times in equipment where the scrubber liquor is in contact with oxygen-containing flue gas; all reactors, mixers, and solids separation equipment should be designed to minimize absorption of oxygen from air. In addition, it has been reported(14) that oxidation of scrubber liquors can be minimized by maintaining very high ionic strength. One possible explanation for this is that high ionic strength liquors are poor oxygen absorbers and that oxidation in these systems is oxygen absorption rate limited.

Water Balance and Waste Product (Cake) Washing - To operate a closed system to avoid potential water pollution problems, system water balance is a primary concern. Water cannot be added to the system at a rate greater than normal water losses from the system.

Generally fresh water is added to a D/A system to serve many purposes, including:

- Saturation of flue gas.
- Pump seal.
- Mist eliminator washing.
- Slurry makeup.
- Waste product washing.
- Tank evaporation.

On the other hand, water should only leave the system through:

- Evaporation by the hot flue gas.
- Water occluded with solid waste product.
- Water of crystallization in solid waste product.

Careful water management, part of which is the use of recycled rather than fresh water wherever possible, is necessary in order to operate a closed system.

As previously indicated, disposal of wet solid waste containing soluble salts is ecologically undesirable. In addition, allowing active alkali or sodium salts to escape from the system is an operating cost factor. Sodium is usually made up to dual alkali systems by adding soda ash (recently quoted at \$60 per ton f.o.b. source) at some point in the system. Thus, both ecological and economic considerations dictate that waste product washing is desirable.

A rotary drum filter, belt filter, or centrifuge is usually where the final solids separation is made. This equipment can be designed for solids washing with fresh water.

One concern in waste product washing is the extent to which the cake should be washed, because of the effect it has on the concentration of solubles in the waste. Solubles in the waste of any FGD system (i.e., lime/limestone and dual alkali) have the potential to contaminate ground water.

One of the major goals of the Resource Conservation and Recovery Act (RCRA)--which requires Federal regulations for disposal of hazardous waste and guidelines (for use by the state) for disposal of non-hazardous waste--is to prevent ground water contamination.

Efforts are currently underway to develop design standards for disposal of utility wastes (i.e., coal ash and scrubber waste) which can be incorporated into RCRA regulations and guidelines.

Although it will be sometime before these standards are developed, an FGD system operator must take this potential for ground water contamination into account in designing his disposal site.

One obvious consideration in waste product washing is system water balance. Unlimited waste product washing is not possible if a closed system operation with no liquid stream discharge is a goal. Another more subtle reason for limiting waste product washing is the potential problem of non-sulfur/calcium solubles buildup in the system. These non-sulfur/calcium solubles enter the system with the fly ash, flue gas, and lime and/or limestone and the makeup water. Of these, probably the soluble material in highest concentration would be sodium chloride which results from the absorption of HCl from the flue gas by the scrubber solution. A material balance around the system at steady state necessitates that solubles leave the system at the rate they enter. Thus, depending upon how well the waste product is washed, a certain level of non-sulfur solubles will be established in the system. Since the only mechanism for these solids to leave the system is as part of the wet solid waste, a certain purge is necessary. This purge also necessitates the loss of some sodium from the system. Practical limitations in filter design and water balance probably would limit a system to two or three "displacement washes" of the waste product (one displacement wash means washing with an amount of fresh water equivalent to the amount of water contained in the final wet waste product per unit of waste). Depending on the characteristics of the waste product and the design of the washing system, one displacement wash can reduce the solubles content of the waste product by as much as 80%.

B.3 ENVIRONMENTALLY ACCEPTABLE SOLID WASTE

Dual alkali systems should be designed to produce an environmentally acceptable end product. Desirable solid waste product properties include:

- Non-toxic.
- Low soluble solids content.

- Low moisture content.
- Non-thixotropic.
- High compressive or bearing strength.

Gypsum vs. Calcium Sulfite - One of the options available to some dual alkali processes is whether or not to oxidize the solid waste. Advantages accruing from oxidation include: gypsum has better handling properties than calcium sulfite because sludges containing a high ratio of gypsum to calcium sulfite are less thixotropic, faster settling, more easily filtered, and can be more completely dewatered than sludges containing a high proportion of calcium sulfite. Another important characteristic which has been attributed to high gypsum (as opposed to calcium sulfite) sludges is their higher compressive or bearing strength.

An explanation for the behavior of high sulfite sludges is given by Selmeczi and Knight.⁽²⁶⁾ Although filter cakes appear dry, they still contain a considerable amount of water and thus, upon vibration or application of stress, have a tendency to again become fluid. This thixotropic property and high moisture content are both explained by the morphology of calcium sulfite clusters. Because of the highly open, porous, or sponge-like nature of these clusters, a considerable amount of water is retained in the clusters. The calcium sulfite crystals are rather fragile and break under pressure, releasing some of the water, which result in fluid sludges.

In Japan, where by-product gypsum is saleable, the calcium sulfite solids produced are oxidized completely to gypsum in a separate oxidation process tacked on to the tail end of the system. In applications where high excess combustion air is present, where low-sulfur coal is burned, or a combination of these conditions, the oxidation rate in the system tends to be high (possibly about 90%) and the proportion of gypsum in the sludge tends to be high. In some dilute systems, the proportion of gypsum in the sludge can be increased by augmental aeration of the scrubbing liquor.⁽⁸⁾ Crystal seeding techniques used in conjunction with augmental aeration can produce relatively coarse grained gypsum crystals with good dewatering and structural properties in the final waste product.

Sludge Fixation Technology - Chemical or physical fixation of the sludge produced in a dual alkali system is another potentially important means of producing an environmentally acceptable solid waste product. This technology is commercially offered by I.U. Conversion Systems, Inc., Dravo Corporation, and Chemfix Corporation. Most of their efforts are concentrated on sludge produced from the more prevalent lime/limestone systems; however, there has been some evaluation of dual alkali sludges. The objective of sludge fixation technology is the production of a non-toxic, unleachable solid waste product which has reasonably high load bearing strength. If dual alkali sludges are amenable to this type of treatment, the need to reduce soluble sulfates in the solid waste product is mitigated. Some sodium sulfate has been found to be physically or chemically tied up in the solid calcium sulfate/sulfite crystal lattice;⁽¹⁴⁾ however, the extent of this phenomenon is not generally considered to be adequate to remove all of the sodium sulfate produced by oxidation. Sludge fixation technology may be a mechanism by which additional sodium sulfate can be removed

from the system without adverse environmental effects. There is some concern that this is not viable, however, since sludge fixation chemistry involves pozzolanic reactions between calcium compounds and fly ash components in the sludge which may only involve multivalent ions rather than monovalent sodium. In other words, monovalent ions such as sodium may either a) not take part in the pozzolanic reactions, or b) inhibit or limit such reactions. Further investigation is needed in this area.

B.4 ECONOMICS

Dilute vs. Concentrated System - The selection of a dilute or concentrated dual alkali system is an important consideration in any application. In general, concentrated systems are desirable for applications where oxidation is expected to be relatively low. Conversely, dilute systems are suitable for applications where oxidation rates are high. For high-sulfur Eastern coal applications on utility boilers (where excess air is controlled carefully and maintained at the lowest value consistent with complete combustion), concentrated systems are favored. On the other hand, in utility or industrial boiler applications (where Western low-sulfur coal is burned, and/or where control of oxidation is difficult due to high excess air), dilute systems may be more suitable.

Oxidation rate is promoted when low-sulfur coal is burned, since the ratio of oxygen to SO_2 in the flue gas is higher than in high-sulfur coal applications. Since oxidation is a strong function of the rate of absorption of oxygen, liquor which is dilute in TOS is subject to having a greater proportion of these species oxidized by a given amount of absorbed oxygen than one in which the TOS is more concentrated.

Under a given set of conditions, without consideration of waste disposal, a concentrated system can be installed at lower capital cost than a dilute system as previously discussed; however, the desirability to produce a manageable solid waste (dilute systems can be designed to produce high gypsum sludges) could, in some cases, override the capital cost issue.

Cost Data - Based on available information, dual alkali systems are economically competitive with the "first generation" wet lime/limestone slurry scrubbing systems. This is particularly true in cases where the lime or limestone system would be required to be equipped with solids separation equipment (i.e., thickener and filter).

Factors that allow a dual alkali system to be less expensive than a lime/limestone system, both in initial cost and annualized operating cost, include:

- Lower scrubber liquid/gas ratio (L/G).
- Lower scrubber pressure drop (ΔP).
- Simpler scrubber design:
 - fewer stages.

- no slurry in scrubber.
- Less exotic materials of construction.
- Solid waste with better handling properties.

It is estimated⁽²⁹⁾ that a new (or simple retrofit) dual alkali system could be installed at a 500 MW plant for a cost around \$100/kW and operated at 4.19 mills/kWhr excluding sludge disposal (which would vary with the particular application). These estimates are based on the following:

- 500 MW or larger system.
- 3.5 wt.% sulfur coal.
- 80% load factor.
- Capital charges @ 15-16% of capital investment, annually.
- Maintenance @ 3-4% of capital investment, annually.
- Operation with two operators/shift.
- Power @ \$0.03/kWhr.
- Soda ash @ \$90/ton (delivered).
- Lime @ \$42/ton (delivered).

THE FGD REAGENT DILEMMA
LIME, LIMESTONE, OR THIOSORBIC LIME

by

Donald H. Stowe
David S. Henzel
David C. Hoffman

Dravo Lime Company
Pittsburgh, Pa.

Introduction

The Clean Air Act of 1970 forced the utility industry into the chemical processing field because of its restrictions on sulfur dioxide (SO_2) emissions. Chemical reactions are required to remove the SO_2 from the combustion of coal. In the early seventies, SO_2 compliance could also be achieved by burning "compliance" coal. This alternative has rapidly disappeared as compliance coal is becoming scarce and expensive. In 1977, the Clean Air Act Amendments were proposed and, if enacted, will make scrubbing an integral part of all new coal-fired electrical power plants.

Since SO_2 removal is a major expense, utilities will attempt to obtain SO_2 compliance at the lowest possible cost; including reliability and maintenance. In order to achieve this, one must realize that these interrelated unit processes--boiler operation, scrubbing, and waste disposal--effect SO_2 removal. Central to the optimization of these three processes is the selection of an alkaline reagent for SO_2 removal. The most prevalent reagents used today are limestone, lime, and Thiosorbic® lime. This discussion will briefly review the basic chemistry involved with each reagent use, outline major advantages and disadvantages of each, and present a realistic example of the cost differences between limestone, lime, and Thiosorbic lime scrubbing.

® - Registered trademark of Dravo Corporation.

Discussion

Generally, scrubbing is usually referred to as the removal of sulfur dioxide (SO_2) from the flue gas of a utility boiler. The source of the SO_2 is the sulfur content in the combustion fuel, most commonly bituminous coal. Current combustion coals may be ranked as low sulfur ($\leq 1\%$), medium sulfur (1-3%), and high sulfur ($\geq 3\%$). Presently, the most generally referred to guideline for SO_2 control is the Federal New Source Performance Standard (NSPS) calling for a maximum 1.2 lb. of SO_2 emission per million BTU's of heat input. Frequently, state and local regulations are even more stringent.

As near term alternatives to FGD, a station may possibly meet regulations by the following:

1. Switching to low sulfur fuel (compliance coal).
2. "Deep" cleaning of the coal prior to combustion.
3. Solvent refining of the coal prior to combustion.
4. Gasification or liquefaction of coal and combustion of resultant gas.
5. Fluidized bed combustion of coal (new boiler design).

Items 3 through 5 have not been demonstrated on a commercial scale although extensive development work is being performed. Of the remaining two alternatives to FGD, fuel switching is the predominant alternative. However, the use of "compliance" coal is being restricted by the following:

1. Availability of product - current sources are the West, Northwest, and sections of Appalachia.

2. Costs associated with production, transportation, and boiler modifications.
3. State regulations prohibiting use of imported coal (i.e., State of Ohio).

In summary, if all the factors are just right, burning of compliance coal may be cheaper than scrubbing. However, this is generally not the case. It has been estimated that compliance coal will be available until 1985 and after that, only high sulfur coal will be obtainable. For the scrubbing industry with a construction lead time of about 3 years, and longer times for new mine-prep plant development; the effect of new, low sulfur, coal sources on the utility market is nearly exhausted if a utility does not already have commitments for the coal. Deep cleaning of coal may be the only near term alternate to FGD. However, the amenability of coals to this process is limited, due primarily to the organic sulfur content of the coal. Consequently, FGD, as we currently know it, will probably be the most widely used technology for the next 10 years.

The non-regenerative lime/limestone processes are the most commonly applied commercial FGD systems in the United States. Primarily, this is due to the lower costs of the system, availability of reagents, ease of operation, system reliability, and experience. Table 1 presents a summary of the total operational commercial FGD systems, as of July, 1978. Of the total 14,420 megawatts, 13,496 megawatts (94%) represent wet calcium-based lime/limestone processes.

Table 1

Operational FGD Systems By Chemical Process - July, 1978

Process	No. of Units	Total MW	Months Experience
<u>Lime/Limestone</u>			
Thiosorbic Lime	6	3,370	151
Lime	4	679	175
Carbide Lime	4	851	93
Lime/Alkaline Fly Ash	3	1,170	65
Limestone	13	6,006	383
Limestone/Alkaline Fly Ash	2	1,420	43
Subtotal	<u>32</u>	<u>13,496</u>	<u>910</u>
<u>Other</u>			
Magnesium Oxide	1	120	34
Sodium Carbonate	3	375	126
Wellman Lord	2	429	23
Subtotal	<u>6</u>	<u>924</u>	<u>183</u>
Total	<u>38</u>	<u>14,420</u>	<u>1,093</u>

Source:
B. Loseke, et.al., EPA Utility FGD Survey: June-July 1978
EPA 600/7-78-051d

Of the total lime/limestone units currently operating, only seven have design removal efficiencies of greater than or equal to 85 percent; the proposed new federal regulation. A breakdown of these units by process is shown in Table 2. Based on this 85% removal criterion, the greatest operational experience lies with Thiosorbic lime.

Process Chemistry

As previously mentioned one of the primary reasons for the dominance of lime/limestone scrubbers is the relative simplicity of the chemical reactions involved. The aqueous sulfur dioxide reacts with the more basic anionic species to produce bisulfite. The predominant cation is calcium. Those basic reactions which occur in these types of systems are listed in Table 3.

Although the lime/limestone processes appear relatively simple, their capabilities are limited by the solubility of the basic anions involved. As a result of this, these processes exhibit two disadvantages: 1) the need for high mechanical energy, and 2) their high potential for gypsum scaling. These systems are basically liquid film limited; that is insufficient dissolved species are present to neutralize the absorbed SO_2 . Therefore, the amount of SO_2 removed is a function of the rate at which the basic species go into solution. Since sufficient dissolved alkalinity is not present, the mechanical energy must be increased to bring more liquid into contact with the gas stream.

Table 2

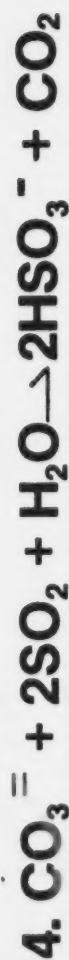
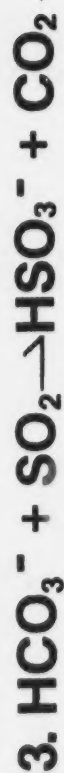
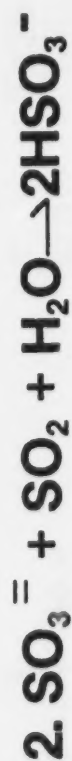
Operational Lime/Limestone FGD Systems Designed for $\geq 85\%$ Removal - July 1978

	Design SO ₂ Removal	MW	Design Coal % S	Months Experience
Thiosorbic				
Mansfield 1	92.1	825	4.7	27
Mansfield 2	92.1	825	4.7	12
Conesville 5	89.5	400	4.7	18
Conesville 6	89.5	400	4.7	1
		<u>2,450</u>		<u>58</u>
Carbide Lime				
Cane Run 4	85.0	178	3.75	23
Cane Run 5	85.0	183	3.75	7
		<u>361</u>		<u>30</u>
Limestone				
Duck Creek	85.0	400	2.75	0
		<u>400</u>		<u>0</u>

Source:
B. Laseke, et.al., EPA Utility FGD Survey: June-July 1978
EPA 600/7-78-051d

Table 3

Basic Reactions Lime/Limestone Systems



Calculations have been performed to determine the theoretically required dissolved alkalinity to achieve 85 percent removal as a function of liquid to gas ratio and inlet sulfur dioxide concentration. The results are tabulated in Table 4. It should be emphasized that these values do not include a factor for the mass transfer capabilities of the scrubber. Typical dissolved alkalinity values obtainable with conventional lime/limestone systems are in the order of 100 to 300 mg/l, as CaCO_3 . Referring to Table 4, it can be seen that these low alkalinities are adequate only at low SO_2 concentrations and/or high liquid to gas ratios.

The introduction of magnesium in the scrubbing loop produces some dramatic changes in the dissolved liquor composition. The solubility of magnesium sulfite is more than two orders of magnitude greater than that of calcium sulfite. As a result of this, an increase in dissolved magnesium produces an increase in sulfite concentration, thus dissolved alkalinity. The final result, of course, is higher removal efficiency. Dissolved sulfite concentrations are generally increased to greater than 2000 mg/l as compared to the typical range of 100 to 200 mg/l obtainable in a conventional lime/limestone system. Alkalinities increase from less than 200 mg/l to greater than 1000 mg/l. Tests have shown that the Thiosorbic process is doing the work of 2.0 to 2.5 identical calcium-based scrubbers. Because of this, it is not only practical to obtain high removal efficiencies (590%) on high sulfur coal, but it also reduces the overall mass transfer characteristics required of the system. Figure 1 exemplifies the net effect of dissolved magnesium upon SO_2 removal.

Table 4

Theoretically Required Dissolved Alkalinity To Achieve 85 Percent Removal

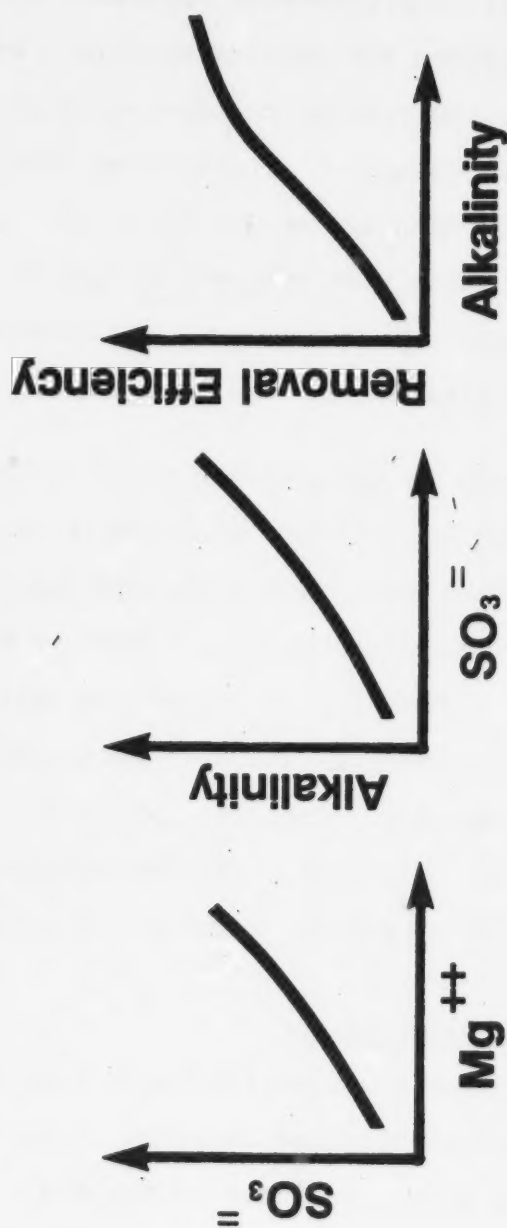
L/G	Inlet SO ₂ (ppm)			
	1,000	2,000	3,000	4,000
30	473	946	1,420	1,892
40	354	710	1,064	1,420
50	284	568	852	1,136
60	236	473	710	946
70	203	405	609	811
80	178	354	532	710
90	158	315	473	631
100	142	284	426	568

Legend:

L/G - Gallons of Scrubbing Liquor per 1,000 SCF

Figure 1

Effect of Magnesium Upon Removal Efficiency



The second key characteristic of magnesium-enhanced scrubbing is the ability to operate subsaturated with respect to gypsum. All lime/limestone processes are based upon operation in a mode saturated for calcium sulfite. With increasing sulfite concentration, the calcium concentration is thereby suppressed. Typical calcium values of greater than 1000 mg/l are readily decreased to less than 100 mg/l. This, of course, reduces the relative saturation with respect to gypsum even though dissolved sulfate concentration increases. This overall effect of magnesium upon gypsum saturation is shown graphically in Figure 2.

In addition to the work performed by Dravo Lime, magnesium-enhanced scrubbing has been extensively studied by the EPA at the Shawnee test facility. Empirical equations have been generated to calculate removal efficiency in a spray tower and relative gypsum saturation as a function of liquor composition. Their equations, along with experimental data have been used to generate the data presented in Table 5. From the table, two trends are evident:

- 1) as magnesium increases, relative saturation decreases, and
- 2) as magnesium increases, removal efficiency increases.

Pilot Plant Demonstrations

Several direct comparisons have been made between lime, limestone, and Thiosorbic lime in pilot plant facilities. Results of one series of tests are shown in Table 6. The 50 cfm pilot, a low energy venturi followed by a spray tower, had a mix tank residence time of 1½ minutes during the lime and Thiosorbic lime

Figure 2

Effect of Magnesium Upon Gypsum Saturation

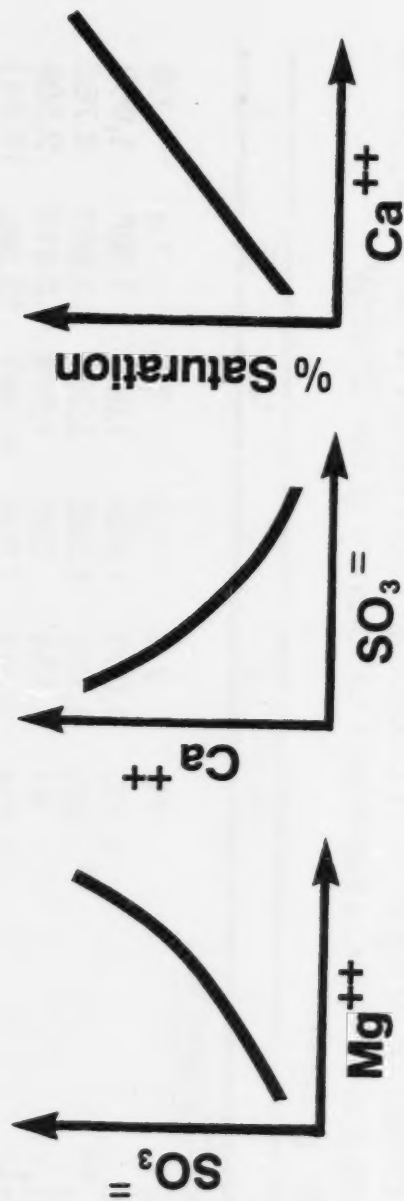


Table 5

The Effect of Magnesium Concentration Upon Relative Gypsum Saturation and SO₂ Removal Efficiency (Low Energy Spray Tower)

	A	B	C	D	E	F
Liquor Composition						
pH	7.0	7.0	7.0	7.0	7.0	7.0
Cl ⁻	1,000	1,000	1,000	1,000	1,000	1,000
Mg ⁺⁺	660	1,180	1,940	3,200	3,960	4,760
SO ₃ ⁼	91	134	285	1,604	2,010	2,908
SO ₄ ⁼	4,704	5,701	7,925	10,867	13,383	15,443
Ca ⁺⁺	920	500	250	60	60	50
% Saturation	154	84	42	10	9	8
Removal Efficiency						
L/G = 40	50.2	55.4	63.2	75.9	82.8	88.9
50	58.2	63.5	71.3	83.1	88.9	93.6
60	64.9	70.2	77.6	88.1	92.8	96.3
70	70.5	75.7	82.6	91.7	95.4	97.9
80	75.2	80.1	86.4	94.2	97.0	98.8
90	79.2	83.7	89.4	95.9	98.1	99.3
100	82.5	86.7	91.8	97.1	98.8	99.6

Table 6

50 CFM Venturi, Spray Tower Comparison of Limestone, Lime and Thiosorbic Lime

Test No.	Reagent	SO ₂ Inlet	L/G Total	Removal Eff.
1	Limestone	1,500	50	59%
2	Limestone	1,500	60	65%
3	Limestone	1,500	90	78%
4	Lime	3,000	60	53%
5	Lime	3,000	80	67%
6	Thiosorbic	3,000	36	57%
7	Thiosorbic	3,000	60	82%
8	Thiosorbic	3,000	82	90%
9	Thiosorbic	3,000	90	93%

tests and 16 minutes for the limestone test. It should be noted the removal efficiencies obtained are somewhat lower than typical values obtained today; however, the trend of increased removal efficiency and lower L/G requirement is very evident.

A two stage 1500 cfm venturi was later operated with an inlet SO_2 concentration of approximately 3000 ppm SO_2 . Tests were conducted with both high calcium lime and magnesium-enriched lime. Results in Table 7 show the increased alkalinity and higher removal efficiency. If we compare theoretical transfer units, it is evident the Thiosrbic lime was doing the equivalent work of approximately two scrubbers in comparison to the high calcium lime scrubber.

Table 8 lists results of tests conducted on a 750 cfm turbulent contact absorber at 3000 ppm inlet SO_2 concentration. Once again, the superiority of the magnesium-enriched system is very evident.

Extensive testing was also performed on a full scale FGD system consisting of single stage venturis. On high calcium lime, typical removal efficiencies ranged from 50 to 65 percent with severe internal scaling, so much in fact that the throat dampers could not be operated. Thiosorbic lime operation not only increased the removal efficiency to greater than 80 percent; but also eliminated the scaling problems; so much in fact that most scale present prior to the Thiosorbic operation actually dissolved.

Table 7

1500 CFM Venturi Comparison of Lime and Thiosorbic Lime

L/G Per Stage	Alkalinity PPM		% Removed		Transfer Units		N_t Thio./ N_t Ca
	High Ca	Thio.	High Ca	Thio.	High Ca	Thio.	
1 40/40	100-140	600	70	92	1.20	2.53	2.1
2 40/40	-	1,100	-	96	-	3.22	-
3 50/40	100-140	600	82	97.2	1.71	3.58	2.1
4 50/50	100-140	600	88	99.6	2.12	5.52	2.6
5 40/--	-	800	-	80	-	1.61	-
6 50/--	-	800	-	87	-	2.04	-

Note:

$$\text{Transfer Units} = N_t = \ln \frac{[\text{inlet SO}_2]}{[\text{outlet SO}_2]}$$

Table 8

750 CFM TCA Comparison of Lime and Thiosorbic Lime

Test Condi- tion	Stage	L/G	SO ₂ Inlet	Removal Efficiency %		Number of Transfer Units		N_t Thio./ N_t Ca
				High Ca	Thio.	High Ca	Thio.	
A	1	20	1,315	60		.92		
	1	40	1,351	86		1.97		
	1	60	1,340	94		2.81		
B	1	30	3,150	45	81	.60	1.66	2.8
	1	40	3,100		84	-	1.83	
	1	50	3,100	59	86	.89	1.97	2.2
	1	60	3,050	72		1.27		
C	2	30	2,850	60	90	.92	2.30	2.5
	2	40	2,850	70	94	1.20	2.81	2.3
	2	50	2,900	79	97	1.56	3.51	2.3
	2	60	2,800	90		2.30		
D	3	20	3,050	60		0.92		
	3	30	3,100	76		1.43		
	3	40	3,100	85		1.90		
	3	50	3,100	90		2.30		

Note:

$$\text{Transfer Units} = N_t = 1n \frac{[\text{inlet SO}_2]}{[\text{outlet SO}_2]}$$

Perhaps the most exemplary comparison of magnesium-enriched scrubbing versus high calcium has been performed by EPA at the Shawnee test facility. Some average liquor compositions are listed in Table 9. The liquor composition of run 619-1A, with high calcium lime, contains a very low sulfite concentration of 40 ppm. A sulfite level of this nature will result in two characteristics, low removal efficiency and gypsum saturated operation. On the other hand, run 611-1A with magnesium-promoted lime and approximately 3200 ppm dissolved magnesium; a sulfite concentration of about 500 ppm results. The higher sulfite concentration not only increased dissolved alkalinity and removal efficiency but at the same time, suppressed the calcium concentration from 1860 ppm to 320 ppm. Tests results show the addition of magnesium not only decreased the relative gypsum saturation from 125 to 45 percent; but at the same time increased removal efficiency from 84 to 95 percent. Figures 3, 4 and 5 are superimposed test results for the two runs. Advantages of the magnesium are very evident.

Full Scale Operations

Six full scale units are currently operating with the Thiosorbic lime process. These units include Mansfield 1 and 2, (1650MW); Conesville 5 and 6, (800MW); Elrama Station (510MW); and Phillips Station (410MW). Sulfur dioxide emission limits are 0.6 lbs/MBTU with the exception of the two units at Conesville which have a limit of 1.2 lbs/MBTU. Sulfur content of the coals range from 2.0 to 4.7 percent, by weight. All of the units currently operating have passed compliance tests, with the exception of Conesville #6. One additional unit, Pleasants #1 (625MW), is in the start-up phase.

Table 9

Shawnee Test Results Average Liquor Composition

	Lime	Lime + MgO
Run	619-1A	611-1A
pH ⁺⁺	7.8	7.0
Ca ⁺⁺	1,860	320
Mg ⁺⁺	330	3,200
SO ₃ ⁼	40	510
SO ₄ ⁼	1,830	11,000
Cl ⁻	2,920	2,500
% Sulfate Saturation	125	45

Figure 3

Comparison of High-Calcium Lime and Magnesium-Containing Lime

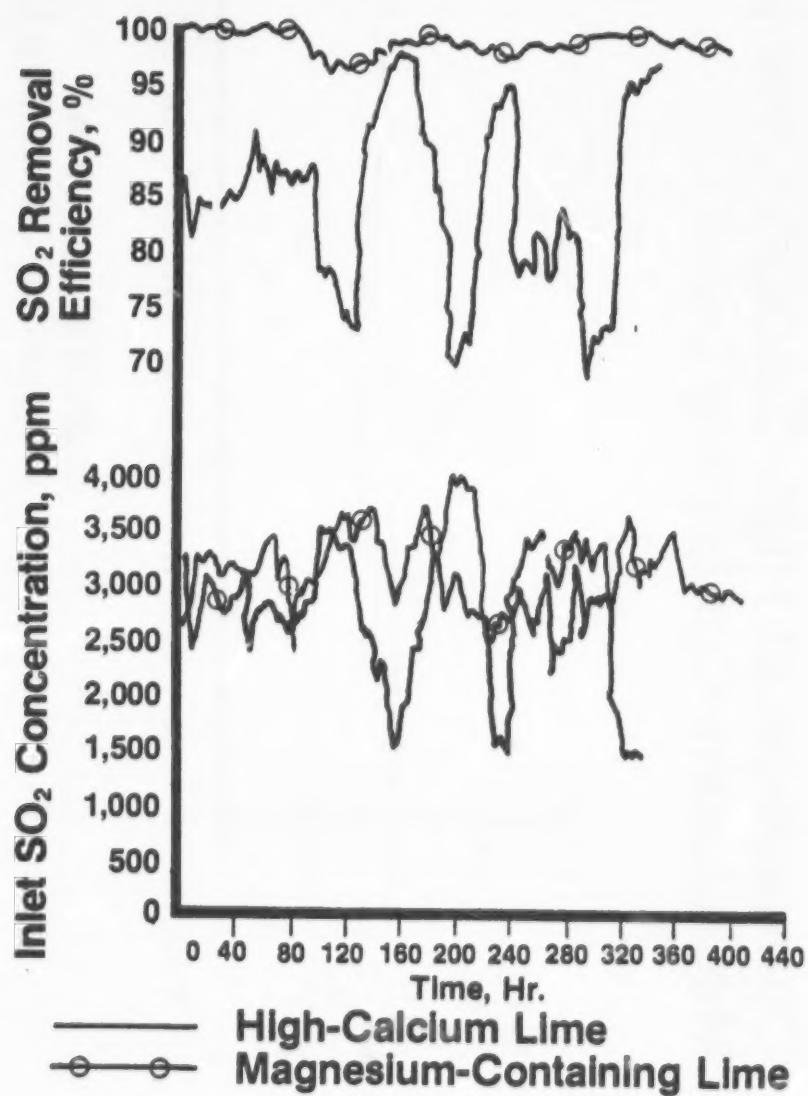


Figure 4

Comparison of High-Calcium Lime and Magnesium-Containing Lime

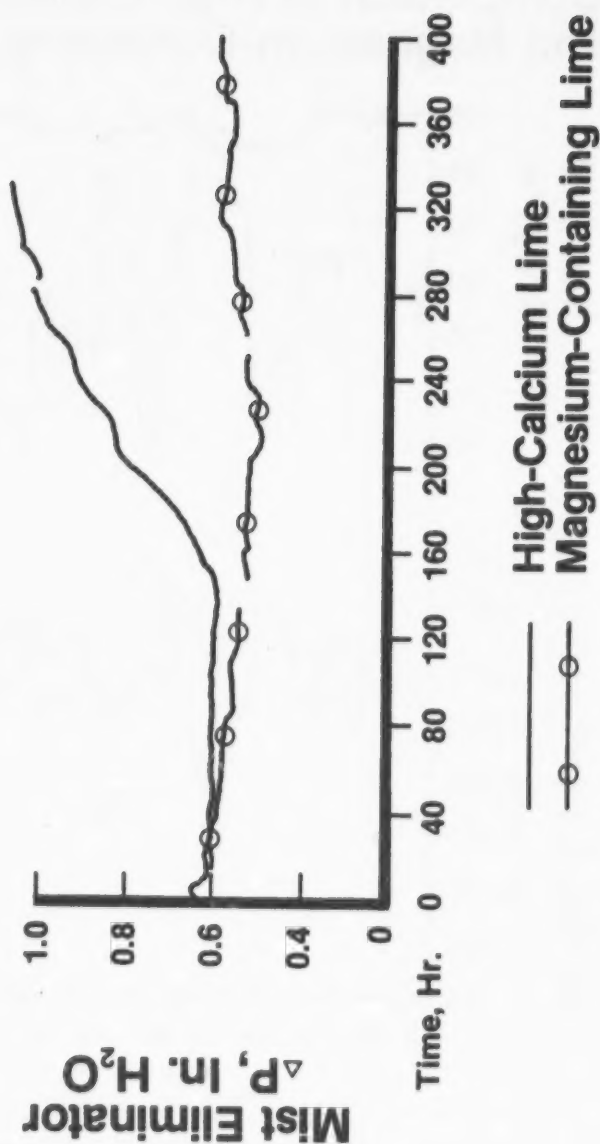
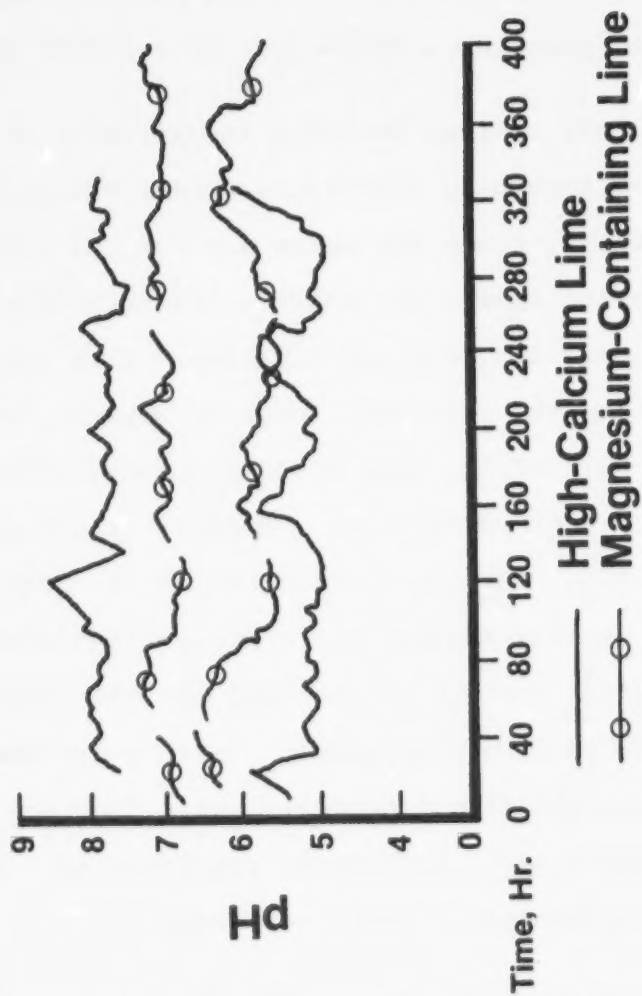


Figure 5

Comparison of High-Calcium Lime and Magnesium-Containing Lime



System Economics

It is extremely difficult to make a direct economic comparison of Thiosorbic lime, lime and limestone scrubbing systems. Perhaps the best available comparison is the work performed by PEDCo under EPA contract. Their work consists of a survey of currently operating FGD systems with adjustments made, thus enabling a comparison on a common basis.

The only systems designed specifically to date for the Thiosorbic scrubbing process have been Pleasants 1 and 2. Unfortunately, since the units are not yet operational, operating costs are not known. All other units currently using our process were designed for high calcium lime. This discussion will therefore compare the lime and limestone systems, keeping in mind that that the bulk of the lime systems operate with Thiosorbic lime. The work performed by PEDCo compares capital costs on a \$/kW basis and operating costs on a mills/kWh basis. The fallacy of this approach is that significant contributing factors such as the design sulfur content of the coal and the required removal efficiency have been excluded. This is extremely important since the capital and operating costs are a function of the amount of SO₂ removed and removal efficiency requirements. Systems removing less SO₂ will undoubtedly have lower costs.

PEDCo's work encompasses 17 scrubber installations currently operating. Calculations have been performed, based upon design criteria, to evaluate capital and operating costs in terms of dollars per ton of SO₂ removed. Data for these stations

are listed in Table 10. From this table it is evident that even on a dollar per ton basis wide fluctuations still exist. Two of the important variables contributing to the wide fluctuations in costs are start-up date and the difference between new and retrofit units.

If we compare only those new units burning high sulfur coal with start-up during the years 1976-1977 impressive correlations exist. Figure 6 exemplifies the capital requirements in terms of dollars per ton of SO_2 removed for the six units falling into this category; three of which are lime systems using Thiosorbic and three limestone. Values for the lime systems range from \$12.96 to \$18.19 per ton of SO_2 removed. The limestone systems range from \$29.46 to \$32.38. Average values for the two processes are \$16.45 and \$30.55, respectively; for lime and limestone.

Annual operating costs are compared on the same basis in Figure 7. Once again the lime process is less expensive. Values for lime scrubbing range from \$232.10 to \$233.14 per ton of SO_2 removed. On the other hand, limestone costs range from \$234.71 to \$331.98. Average values for the annual operating costs are \$232.70 and \$285.54; respectively, for lime and limestone. Capital and operating costs for a system designed specifically for Thiosorbic scrubbing would yield even higher savings.

Table 10

Operational FGD Systems Economics

Alkali	Unit	FGD MW	New or Retrofit	Startup	Coal % Sulfur	Design SO ₂ Removal	Capital Cost (\$/kw)	Annual Operating Cost (mills/kwh)	Tons SO ₂ Removed (Tons/Yr)	Capital Cost (\$/Ton SO ₂ Removed)	Annual Operating Cost (\$/Ton SO ₂ Removed)
Thiosorbic Lime											
Conesville 5		411	N	1/77	4.7	89.5	70.8	7.42	74,816	12.96	232.10
Elrama 1-4		510	R	10/75	2.0	83.0	134.5	8.59	36,636	62.41	680.88
Phillips 1-6		410	R	7/73	2.0	83.0	149.7	9.54	29,453	69.46	756.18
Mansfield 1		917	N	4/76	4.7	92.1	102.2	7.67	171,774	18.19	233.14
Mansfield 2		917	N	7/77	4.7	92.1	102.2	7.67	171,774	18.19	233.14
Lime											
Hawthorn 3		110	R	11/72	2.0	70.0	87.3	4.09	6,665	48.03	384.36
Hawthorn 4		110	R	8/72	2.0	70.0	87.3	4.09	6,665	48.03	384.36
Green River 1-3		64	R	9/75	3.8	80.0	77.6	5.24	8,419	19.66	226.80
Cane Run 4		190	R	8/76	3.75	85.0	80.6	8.64	26,208	19.48	356.66
Cane Run 5		200	R	12/77	3.75	85.0	67.5	5.56	27,587	16.31	229.51
Paddys Run 6		70	R	4/73	3.75	80.0	76.5	6.51	9,088	19.64	285.53
Limestone											
Cholla 1		126	R	10/73	.55	58.5	56.0	2.58	1,754	139.04	1,055.09
Petersburg 3		532	N	10/77	3.25	80.0	100.6	6.56	59,857	29.80	331.98
LaCygne 1		874	N	2/73	5.0	76.0	68.0	3.78	143,723	13.78	130.89
Winah 2		280	N	7/77	1.0	69.0	66.5	2.92	8,360	74.24	556.87
Southwest 1		194	N	4/77	3.5	80.0	117.7	6.17	23,507	32.38	289.94
Widows Creek 8		550	R	5/77	3.7	80.0	113.2	5.28	70,451	29.46	234.71

Source:

B. Laseke, et.al., EPA Utility FGD Survey: June-July 1978
EPA 600/7-78-051d

Figure 6

**Capital Costs - Dollars per Ton SO₂
Removed New Units Burning >3%
Sulfur Start-Up in 1976-1977**

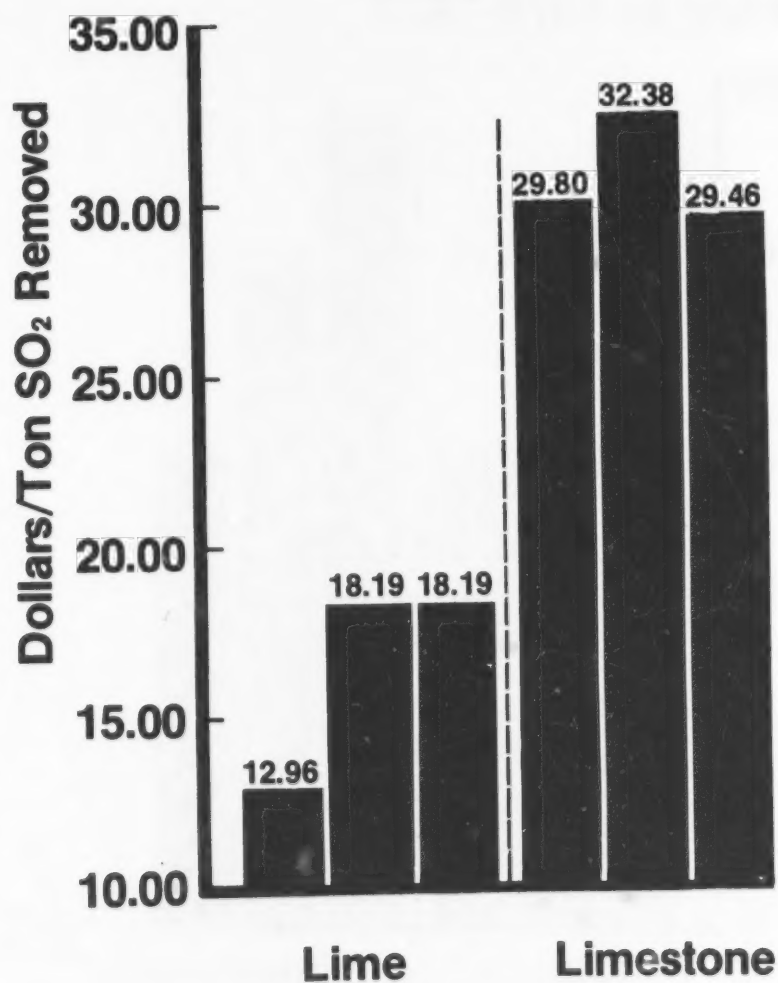
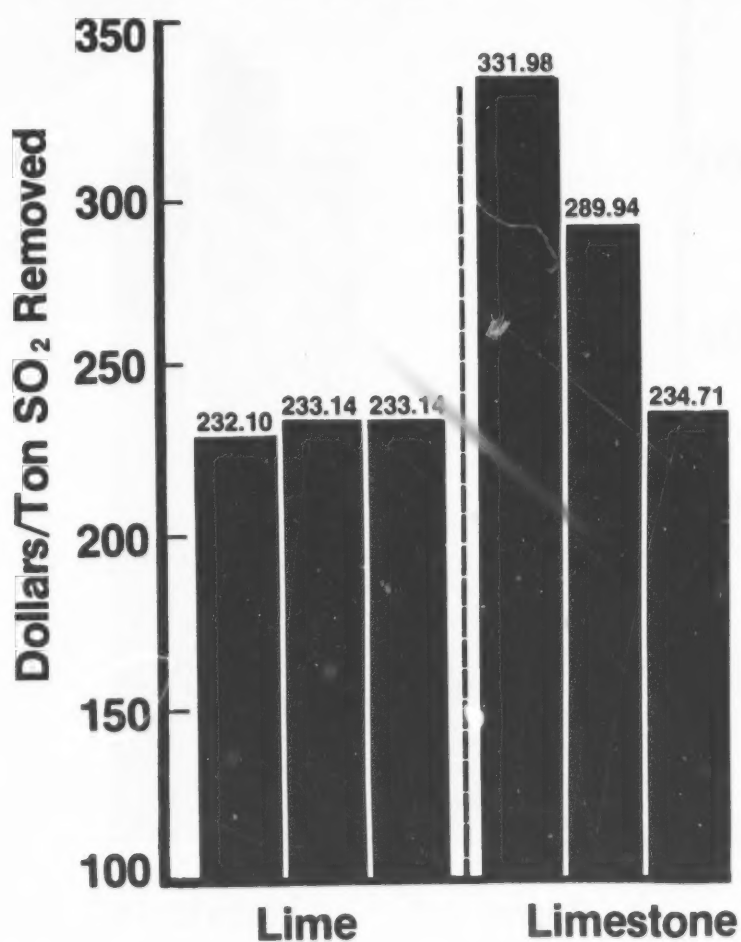


Figure 7

**Annual Operating Costs - Dollars
per Ton SO₂ Removed New Units
Burning >3% Sulfur
Start-Up In 1976-1977**



Conclusion

The discussion today has been directed at reagent decision-making from an operational point of view for those faced with burning high sulfur coal and meeting the proposed SO_2 emission standards. This information is supported by the largest base of experience to date, the Thiosorbic lime process.

Over the last several years, the state of the art within the FGD industry has moved forward at a rapid pace. This success must be attributed to organizations such as EPA, TVA, EPRI, the utilities and private enterprise. Even with the great advances made to date, the magnesium-enriched lime scrubbing system still remains a very favorable option, both technically and economically.

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3. Harlan Head, "EPA Alkali Scrubbing Test Facility: Advanced Program", Second Progress Report. EPA 600/7-76-008, September, 1976.
4. Harlan Head, "EPA Alkali Scrubbing Test Facility: Advanced Program", Third Progress Report. EPA 600/7-77-105, September, 1977.
5. B. Laseke, et.al., "EPA Utility FGD Survey", EPA 600/7-78-051d, June-July, 1978.
6. Federal Power Commission, "The Status of Flue Gas Desulfurization Applications in the United States: A Technological Assessment", NTIS PB-271-362, July, 1977.

SESSION 5

**FGD CURRENT STATUS AND FUTURE PROSPECTS:
VENDOR PERSPECTIVES**

FRANK T. PRINCIOTTA, CHAIRMAN

Panel: Brief discussion of significant issues
followed by questions from the audience.

Members: Abdus Saleem
Irwin A. Raben
James R. Martin
Henry M. Majdeski
Robert J. Gleason
Vincent B. Birkner

No papers or discussions are included for
this session.

SESSION 6
INDUSTRIAL APPLICATIONS
RICHARD D. STERN, CHAIRMAN

THE STATUS OF INDUSTRIAL BOILER
FGD APPLICATIONS IN THE
UNITED STATES

Prepared by:

John Tuttle
Avinash Patkar
PEDCo Environmental, Inc.

and

R. Michael McAdams
U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina

For Presentation at the U.S. EPA
Las Vegas FGD Symposium

March 5-8, 1979

ABSTRACT

PEDCo Environmental, Inc., under contract to the U.S. EPA's Industrial Environmental Research Laboratory at Research Triangle Park, has been monitoring the status of industrial boiler flue gas desulfurization (FGD) applications since 1976. The information provided in this survey program has been obtained by visits to these industrial FGD sites, and through regular contact with company representatives, process designers, equipment suppliers, and government agencies.

This paper summarizes the EPA Industrial Boiler FGD Survey report, including:

- ° The current (fourth quarter 1978) status of industrial FGD applications in the U.S., identifying the number of systems operating, under construction, or in a planning phase.
- ° A summary of system suppliers, including the number of systems served, processes offered, and total gas flow treated.
- ° A summary of FGD process types currently being implemented, with a discussion of some major installations, including system design and operating experience.
- ° A summary of currently reported costs for the industrial FGD applications.

INTRODUCTION

This paper addresses industrial boiler flue gas desulfurization (FGD) applications, a topic about which little was widely known until rather recently. Under contract to U.S. EPA's Industrial Environmental Research Laboratory at Research Triangle Park in North Carolina, PEDCo Environmental has been monitoring FGD technology as applied to industrial boilers since the summer of 1976. The product of this program is the quarterly updated report titled, "EPA Industrial Boiler FGD Survey." Anyone desiring to receive this report, at no cost, may request the addition of his or her name to the mailing list by contacting one of the authors of this paper.

Heretofore much of the focus of FGD has been on utility applications. This has been due to the high levels of emissions from large utility plants, the tremendous capital and operating costs associated with utility applications, the need for rapid FGD development in the utility sector, and the good communication among the various utility companies. During the past year or so, however, increased attention has been given to the problem of sulfur dioxide (SO_2) emissions from industrial boilers.

Therefore, a survey was instituted to summarize the status of FGD systems in the industrial sector. In this survey effort no claim is made that all systems operating, under construction, or planned are identified. Indeed, several known FGD systems are

not included in the report, at the request of the FGD plant representatives. However, we do maintain close contact with these plants and monitor their FGD activities. In addition, there are other known FGD systems whose company representatives choose to maintain confidentiality concerning their activities. Despite these deficiencies, the survey report presents a fairly representative picture of FGD systems on industrial boilers.

In matters concerning utility companies, there is a well-developed communication network. Items such as rates, fuels, load forecasts, boiler sales, and pollution abatement strategies are well publicized and addressed in many arenas. The activities of the private corporations, though not necessarily secret, are simply not as highly publicized in such a coordinated manner.

Basic differences exist between utility and industrial applications. Some of these differences are discussed in the following paragraphs.

The most obvious difference is size. Whereas the average utility FGD installation is on the order of 350 MW (about 700,000 scfm)*, the average industrial application is on the order of 40,000 scfm (about 20 MW equivalent). However, in comparing numbers of systems, there are 46 operating utility systems¹ and 132 operating industrial systems.²

Another basic difference is the type of technology being used to treat the flue gas in the two different applications. In

* The British measurement system is used in this report, despite EPA's metric policy, for convenience to the readers. The British to metric conversion factors are given at the end of this paper.

utility applications, about 90 percent are treated by calcium-based technology (lime or limestone). In industrial applications, about 90 percent are treated by sodium-based technology (sodium hydroxide, sodium carbonate, dual alkali, or caustic waste stream). This affects the equipment required, types of problems encountered, control level attainable, and the types of end materials generated.

Yet another difference is that, in general, the smaller industrial FGD installations incorporate less redundancy, less ancillary equipment, and less instrumentation. However, operating histories have been generally successful. The most recent survey report, for the fourth quarter of 1978, listed no fewer than 18 sites reporting 90 percent or better reliability. These sites represent 90 FGD systems controlling 109 boilers or steam generators. Nine of these sites reported a 100 percent reliability index value for their 14 FGD systems (controlling 32 boilers or steam generators). Three other sites did not report the relevant operating hours although they did report that no problems occurred. Of the remaining 18 operating sites, operating hours were made available by only 2 sites; GM's Parma FGD system demonstrated a 46 percent reliability for October and November, and GM's Dayton (Delco Moraine) system was down for the whole period.

The fourth quarter of 1978 was not unique with respect to these high system reliabilities. It has become clear that the

application of FGD technology in the industrial sector has been generally successful.

Table 1 summarizes the number and gas flow capacity of FGD systems operating on industrial boilers in the United States. It can be seen that the number of operating industrial FGD systems (132 as of December 1978) far exceeds operating utility FGD systems (46 as of January 1979). However, utility FGD applications are far exceeding industrial applications with respect to the amount of flue gas being treated. Utility applications are treating about 16,000 MW whereas industrial applications are treating 2700 MW equivalent, using a conversion of 2000 scfm per MW equivalent.

Figure 1 shows the steady increase in the number of FGD applications to industrial boilers beginning in 1972. The growth trend is shown to drop off in the 1980's because companies are not making firm plans that far ahead.

Table 2, a breakdown by FGD process and gas flow capacity being controlled, shows that about 86 percent are using a sodium-based SO_2 absorption mechanism. The advantage of sodium-based systems is that absorption is effected by soluble salts. This eliminates the scaling and plugging problems that accompany calcium-based systems.

Tables 3 and 4 summarize the vendors who are serving the industrial FGD market.

TABLE 1. NUMBER AND CAPACITY OF
U.S. INDUSTRIAL BOILER FGD SYSTEMS*

Status	Number of FGD units (sites)	Capacity, scfm
Active		
Operational	132 (39)	5,487,000
Under construction	27 (11)	1,236,000
Planned	14 (11)	2,386,000
Total	173 (61)	9,109,000

* There are probably several systems in various planning phases which have not yet been located.

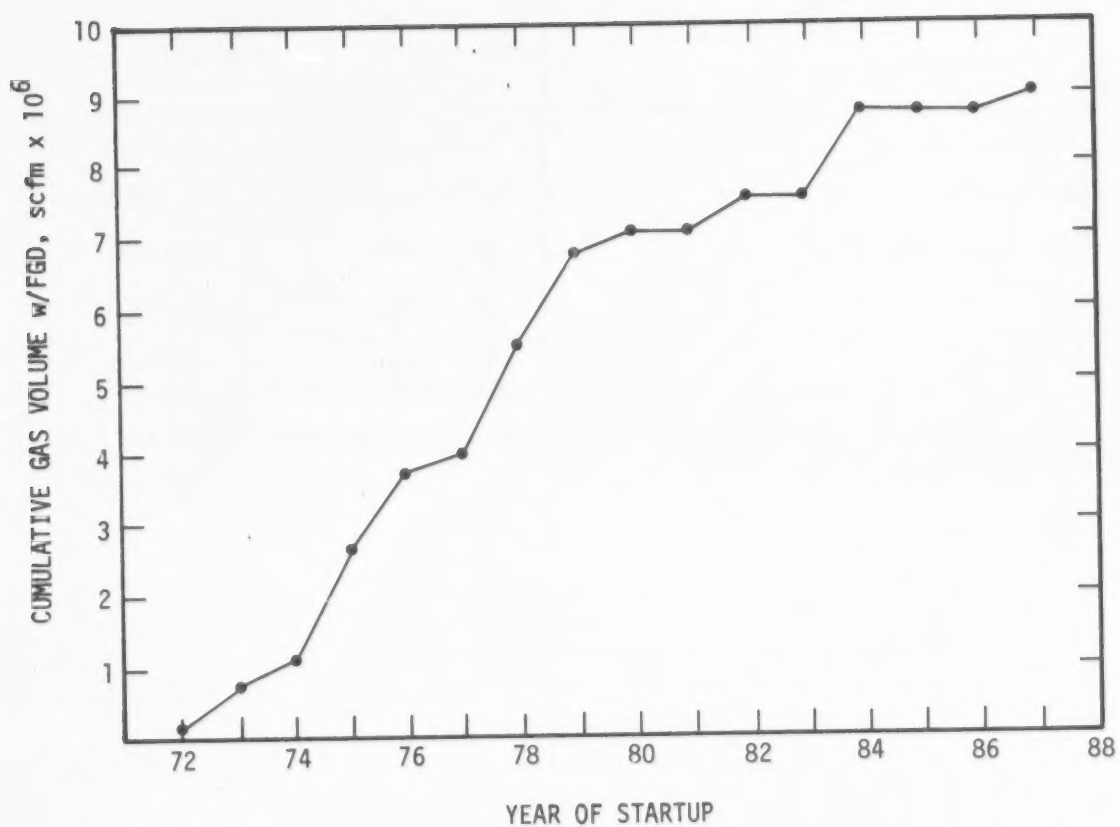


Figure 1. Use of industrial FGD as a function of startup date.

TABLE 2. INDUSTRIAL FGD PROCESS APPLICATIONS

Process	Capacity, scfm x 10 ³		
	Operating	Under construction	Planned
Nonregenerable sodium (sodium hydroxide or sodium carbonate)	3,135	656	286
Dual alkali	591	437	417
Lime	84	0	30
Limestone	55	0	320
Caustic waste stream	1,009	0	0
Ammonia waste stream	626	0	0
Citrate	0	142	0
SULF-X	0	0	10
Process not selected			1,324

TABLE 3. VENDORS WITH MORE THAN ONE INDUSTRIAL FGD APPLICATION

Vendor	No. of installations	Process	Total size, scfm	Status*
Airpol	7	Sodium hydroxide Sodium carbonate Dual alkali Caustic waste stream	1,274,000	O,C,P
C-E Natco	3	Sodium hydroxide	24,000	O,C
FMC Environmental Equipment	12	Sodium hydroxide Sodium carbonate Dual alkali Caustic waste stream	1,585,000	O,C,P
GM Environmental	2	Sodium hydroxide Dual alkali	235,000	O
Heater Technology	2	Sodium hydroxide Sodium carbonate	92,000	O,C
Koch	3	Sodium carbonate	478,000	O,P
Zurn	2	Dual alkali	105,000	O

* Operating, Construction, Planned are O, C, P, respectively.

TABLE 4. VENDORS WITH ONE INDUSTRIAL FGD APPLICATION

Vendor	Process	Size, scfm	Status *
A.D. Little	Sodium hydroxide	64,000	O
Bureau of Mines	Citrate	142,000	C
Carborundum	Lime	30,000	P
Ceilmote	Sodium hydroxide	380,000	O
Combustion Equipment Assoc.	Sodium carbonate	490,000	O
Ducon	Sodium carbonate	117,000	C
Entoleter	Sodium hydroxide	36,000	O
Flakt	Sodium carbonate	39,000	C
Pittsburgh Environmental and Energy Systems	SULF-X	10,000	P
Research-Cottrell/Bahco	Limestone	55,000	O
Swemco	Sodium carbonate	140,000	O
Thermotics	Sodium hydroxide	12,000	O
W.W. Sly Manufacturing	Caustic waste stream	18,500	O
Wheelabrator-Frye/Rockwell International	Dry sodium carbonate	52,300	P

* O = operating, C = construction, and P = planned.

NONREGENERABLE SODIUM FGD TECHNOLOGY

The predominant technology currently in use in industrial FGD applications employs either sodium hydroxide or sodium carbonate as the makeup reagent. Sizes range from 12,000 to 245,000 scfm.

There are 93 operating FGD systems involved at 18 plant sites. Table 5 summarizes these plants. A brief discussion of FGD sites follows.

Alyeska Pipeline Service Company, Valdez, Alaska

Background--

The SO₂ control system at this installation is not specifically an environmental control device, as it is not required to comply with an emissions regulation. The fuel being burned in this three-unit boiler plant is low sulfur oil (0.03 to 0.1 percent). Alyeska needed an inert, noncorrosive blanket gas for its oil storage tanks. It was determined that boiler flue gas would have the required low oxygen levels. However, the SO₂ concentration in the boiler flue gas caused corrosion problems. FMC Environmental Equipment Division solved the corrosion problem by providing a twin-module FGD system incorporating FMC's disc-and-donut tray design (four trays per module). The system is designed with 100 percent redundancy in that only one module is used at a time. The spare module assures Alyeska that it will have an uninterrupted supply of inert gas for use in the oil storage tanks. Scrubber bleed-off (10 gpm) is aerated to sodium

TABLE 5. INDUSTRIAL SITES USING NONREGENERABLE SODIUM FGD TECHNOLOGY

Company	Location	Fuel	Flow rate, scfm	Reagent	Vendor	% SO ₂ removal
Alyeska Pipeline Service Co.	Valdez, AK	Oil	50,000	Sodium hydroxide	FMC Env. Equipment	96+
Belridge Oil Co.	McKittrick, CA	Oil	12,000	Sodium hydroxide	Heater Technology	90
Belridge Oil Co.	McKittrick, CA	Oil	12,000	Sodium hydroxide	Thermotics, Inc.	90
Chevron U.S.A., Inc.	Bakersfield, CA	Oil	248,000	Sodium carbonate	Koch Engineering	90
FMC (Soda Ash Plant)	Green River, WY	Coal	446,000	Sodium carbonate	FMC Env. Equipment	95
General Motors Corp.	Dayton, OH	Coal	36,000	Sodium hydroxide	Entoleter, Inc.	86
General Motors Corp.	Pontiac, MI	Coal	107,300	Sodium hydroxide	GM Environmental	N.A.
General Motors Corp.	St. Louis, MO	Coal	64,000	Sodium hydroxide	A. D. Little	90+
General Motors Corp.	Tonawanda, NY	Coal	92,000	Sodium hydroxide	FMC Env. Equipment	90-95
Getty Oil Co.	Bakersfield, CA	Oil	72,000	Sodium carbonate	FMC Env. Equipment	90+
Getty Oil Co.	Bakersfield, CA	Oil	95,000	Sodium carbonate	In-house Design	96
ITT Rayonier, Inc.	Fernandina Beach, FL	Bark & Oil	176,000	Sodium hydroxide	Airpol Industries	80-85
Kerr-McGee Chemical Corp.	Trona, CA	Coke, Coal, & Oil	490,000	Sodium carbonate	Combustion Equipment Assoc.	98+
Mead Paperboard Co.	Stephenson, AL	Oil	100,000	Sodium carbonate	Airpol Industries	95
Mobil Oil Co.	San Ardo, CA	Oil	175,000	Sodium hydroxide	In-house design	90
Northern Ohio Sugar Co.	Freemont, OH	Coal	40,000	Sodium hydroxide	In-house design	N.A.
Texaco, Inc.	San Ardo, CA	Oil	380,000	Sodium hydroxide	Cellcote	73
Texasgulf	Granger, WY	Coal	140,000	Sodium hydroxide	Swenco, Inc.	90

N.A. = Not available.

sulfate, and mixed with ballast water (10,000 gpm) from tankers in Port Valdez, and is discharged into the bay.

Operating Experience--

Because the scrubber plant is required for successful operations elsewhere on-site, Alyeska spares no expense in operating the FGD system. Although the design control pH was 6.5, control pH is now 8.0, for additional SO_x control. Another unique approach has been the installation of cyclohexylamine sprays downstream of the absorber to reduce corrosion. An emissions test run in late 1977 showed that, while inlet SO_2 concentrations were ranging from 150 to 160 ppm, outlet SO_x ($\text{SO}_2/\text{SO}_3/\text{SO}_4$) levels were about 5 ppm giving approximately 97 percent SO_2 removal efficiency.

Alyeska has released no capital or operating cost data for this control system.

FMC Chemical Corporation Soda Ash Plant, Green River, Wyoming

Background--

This soda ash plant operates two coal-fired boilers firing a 1.0 percent sulfur coal. Each boiler produces 330,000 acfm at 320°F. Particulate control occurs in a hotside ESP, which is followed in sequence by the economizer, a forced draft (with respect to the FGD system) fan, and an FMC Environmental Equipment Division SO_2 control system.

Each boiler has its own FGD system consisting of two disc-and-donut absorber modules. The absorber shells and discs are

carbon steel with a Ceilcote liner; the outer donut portions are Inconel 825. Recirculation pumps and piping are rubber-lined.

To meet applicable codes, approximately 50 percent of the boiler flue gas is scrubbed; the remainder is bypassed and used for reheat. A bleed off stream is taken to a holding pond for evaporation.

FMC reported that the overall air pollution control system cost was \$10 million in 1975 dollars.

Operating Experience--

Tests in 1978 showed that SO_2 removal efficiency varied from 87 to 94 percent depending on L/G ratio and pH. Problems have included: a module liner failure, inadequate cold weather protection, inadequate pH control, faulty damper operation, and a broken shaft on a recirculation pump. These problems have been rectified and the FGD reliability index for the fourth quarter of 1978 was reported at close to 100 percent.

General Motors Corporation's Delco Moraine plant, Dayton, Ohio

Background--

This plant has two coal-fired (0.7-2.0 percent sulfur) boilers each of which generates 34,000 acfm of flue gas at 500°F. Initial particulate collection occurs in internal multiclones; an Entoleter, Inc. emission control system provides secondary particulate and primary SO_2 control. Entoleter's "vane-cage" system sets up a vortex of scrubbing liquor mist. A bleed-off from the recirculation line is pumped to the onsite wastewater treatment

facility, where it is clarified and pH adjusted, then discharged to city sewers.

The design includes 316L stainless steel for absorber shells and internals, mist eliminators, stacks, and recycle tanks. The scrubber forced-draft fans, caustic tank, caustic pump, and inlet ductwork are carbon steel. The recirculation pumps and piping are rubber-lined.

The total installed capital cost was \$668,000 in 1974 dollars.

Operating Experience--

Plant personnel are currently carrying out extensive modifications necessitated by serious corrosion of the primary mist eliminator. The corrosion was caused by acid rain fallout from the secondary mist eliminator (no wash was provided). The new configuration will consist of Hastelloy-G primary and secondary radial vane mist eliminators and an intermediate Hastelloy-G convex impingement plate to assist in rapid drainage of the mist eliminator runoff. New flue gas flow sensors and a new pH monitoring system are also being installed.

Kerr-McGee Chemical Corporation, Trona, California

Background--

In June 1978, production began at Kerr-McGee's new soda ash plant, the largest yet built to yield soda ash by direct carbonation of brine. Annual production will be 1.3 million tons per year. Processing innovations include carbonation under 13.5 psig pressure and the recovery of CO₂ from fossil-fuel-fired boilers.

The plant has two boilers, each producing 600,000 lb/h steam at 1500 psig. The steam is used initially to drive two 32-MW non-condensing steam turbines to generate electricity, and is then used as process steam. The boilers burn a mixture of western coal (0.7 percent sulfur) and petroleum coke (5.5 percent sulfur). The flue gas flow rate from each boiler is 363,000 acfm at 320°F and may contain SO₂ concentrations ranging from 335 to 5985 ppm depending on the fuel mix used.

The flue gases are scrubbed by the end liquor from the soda ash plant in flaked-glass lined mild steel vessels, each with three 317L stainless steel sieve trays. Combustion Equipment Associates supplied the FGD system. Spent liquor from the scrubber bottom is recirculated by rubber-lined pumps; fresh end liquor is added directly to the recirculation line. A bleed stream returns absorber reaction products to the salt ponds.

The scrubbed flue gases are processed through two monoethanolamine plants for CO₂ extraction. Ambient air is heated externally by a steam-tube bank and mixed with exit flue gases for 50°F reheat.

The capital cost of this system was reported to be \$6 million in 1978 dollars.

Operating Experience--

The currently operating 317L stainless steel sieve trays replaced the original polypropylene trays. The holes in the original trays were cut unevenly and were not large enough. No

other problems have been reported. Both scrubbers demonstrated 100 percent reliability for the fourth quarter of 1978.

Mead Paperboard Company, Stevenson, Alabama

Background--

This plant is a 500 ton per day neutral sulfite pulp mill. Two oil-fired (1.5 to 3.0 percent sulfur) boilers produce 175,000 lb/h of steam each, at 600 psig. Airpol Industries supplied the FGD system which treats 173,000 acfm at 450°F. The flue gas passes through a stainless steel venturi/quench section, and then into the 316L stainless steel absorber which includes three bubble-cap trays for SO₂ removal. Sodium carbonate solution is added directly to the recirculation line. Scrubber effluent is sent to the onsite pulping operation by a continuous bleed-off.

Operating Experience--

Early operations were marked by the failure of a flaked-glass liner (stress cracking), and a Carpenter 20 liner (seepage between the liner and the shell). In 1977, a rubber liner was installed throughout the entire scrubber, including the spin vanes. As of the fourth quarter of 1978, the rubber lining was still serving well, and the FGD system demonstrated 100 percent reliability.

Texasgulf, Inc., Granger, Wyoming

Background--

At this 1 million ton per year soda ash plant, the boiler plant consists of two coal-fired (0.75 percent sulfur) boilers

which generate a total of 600,000 lb/h of steam at 300 psig. Primary particulate control occurs in a hot-side ESP SO₂ control occurring in a Swemco, Inc. FGD system which consists of a quencher, a two-stage sieve tray absorber, and a mesh-type mist eliminator. The system is provided with bypass.

The quench section is Inconel 625, the absorber is carbon steel with a flaked-glass liner, the sieve trays are Inconel 625, and the mist eliminator is Teflon. The stack is insulated steel with a flaked-glass liner.

Each absorber is divided vertically into two sections one of which has a damper at the top. The damper allows for turndown to 25 percent of full boiler load. A bleed-off stream goes to a holding pond for evaporation.

Operating Experience--

Only minor problems have been reported since the September 1976 startup. Ductwork at the outlet to the absorber was replaced by 304 stainless steel, some minor nozzle plugging has occurred, and some piping was replaced. Most replacements have been made during scheduled boiler outages. The fourth quarter 1978 reliability index was 100 percent.

California Enhanced Oil Recovery (EOR) Sites

Of the more than 170 FGD units currently covered by the survey, more than half are in use or scheduled for use in EOR sites in California. Since these units predominate in the survey, a brief discussion of the EOR industry follows.

In 1976, the San Joaquin Valley Air Basin produced about 45 percent of California's crude oil.³ EOR operations accounted for about 50 percent of that production. Additionally, EOR operations in the basin accounted for approximately 80 percent of California's total EOR operations.

The most common means of producing crude oil relies on natural underground pressure to push the oil to the surface of a well. If natural pressure is lacking, pumps are used to pull the oil out of its reservoir. These methods of producing crude oil are termed "primary recovery." If water is pumped into an oil reservoir (secondary recovery), crude recovery can be increased or "enhanced." When the capabilities of the primary and secondary methods are exhausted, additional enhanced recovery (tertiary recovery) methods are considered. Tertiary oil recovery employs thermal means to make heavy, viscous crude oils move through sand and rock strata more easily.

Typically steam is injected continuously to flood the underground oil reservoir with steam and hot water. There are currently about 800 steam generators (typical steam quality is 80 percent) in the San Joaquin Basin, most of which have capacities in the range of 15 to 65 million Btu/h. Steam generators in this size range vary from about 5000 to 30,000 acfm at approximately 600°F.

The California Air Resources Board (CARB) has proposed SO₂/NO_x regulations which would require the following for oil-fired steam generators:

<u>Source</u>	<u>SO₂ emission limit, ppm</u>	<u>NO_x emission limit, ppm</u>
New generators	60	100
Existing generators	200	150

To date, the following EOR production companies have given permission to discuss FGD operations at their oil sites:

Belridge Oil Company
 Chanslor Western Oil and Development Company
 Chevron U.S.A., Inc.
 Getty Oil Company
 Mobil Oil Company
 Shell Oil Company
 Sun Production Company
 Texaco, Inc.

These companies are firing oil (1.0 to 1.7 percent sulfur) recovered from their own onsite wells. Furthermore, all are using sodium-based chemistry for SO₂ removal, in scrubber/absorbers supplied by companies such as Thermotics, Heater Technology, Ducon, and C-E Natco, in addition to such firms as FMC Environmental Equipment Division and Koch. Getty Oil and Mobil Oil are designing their own equipment.

At the Chanslor Western site near Bakersfield, California, dual alkali technology will be used to regenerate sodium hydroxide scrubbing liquor and produce a calcium sulfite/sulfate sludge. All of these systems are designed for approximately 90 percent SO₂ removal.

Belridge Oil currently has two operating FGD systems supplied by different vendors, each employing an eductor-type venturi for SO₂ removal. The venturi has an adjustable disc, which forms the contact stage for the gas and the recirculating liquor. L/G ratio is 40 gal./1000 acf. Because gas and liquid move in the same direction and only one stage of absorption exists, a higher pH (about 8.0) is maintained in the recirculating loop. Problems have included a pump failure, pump vibration, and faulty electric hookups. Nevertheless, both units demonstrated better than 90 percent reliability in the fourth quarter of 1978.

Chevron U.S.A., Inc. employs three Koch Engineering FGD systems to control SO₂ emissions from 18 steam generators (six generators per FGD system). The three systems are controlling a total of 450,000 acfm at 500°F (150,000 acfm per system). The scrubber modules consist of a quench section followed by three Koch Flexitrays and a Fleximesh mist eliminator. The scrubbers and trays are 316L stainless steel, the Fleximesh is Incoloy 825, and the stub stack is 316 stainless steel. Recent operations have been completely problem-free.

Getty Oil Company operates an oil field near Bakersfield which occupies 15 square miles. The EOR operations use 136 steam generators, 100 with a heat rate of 50 million Btu/h and 36 with a heat rate of 20 million Btu/h, all firing a 1.1 percent sulfur oil.

Currently six FGD systems are operating to control SO_2 emissions from 51 steam generators. FMC Environmental Equipment supplied a three-stage disc-and-donut (316L stainless steel) absorber followed by an Inconel wire mesh mist eliminator. This system controls emissions from six steam generators each rated at 50 million Btu/h. These systems are briefly discussed as follows.

The FMC system has required some fairly extensive rework including the addition of a fourth absorption tray and a switch-over from a mesh to a chevron mist eliminator. Getty is using an in-house-design which will consist of a system controlling emissions from nine steam generators, each rated at 50 million Btu/h. Each absorber will consist of three Koch Flexitrays and a Flexi-chevron mist eliminator. These systems are designed to remove 96 percent of the inlet 600 ppm SO_2 . As of January 1979, five of these systems were on-line with four more scheduled to start up in mid-February 1979.

The Getty in-house-designed systems have experienced some scaling, vibrating fans, and some poor mist elimination characteristics. Getty will evaluate the costs of running sodium hydroxide as compared to sodium carbonate.

DUAL ALKALI FGD TECHNOLOGY

In industrial FGD applications, dual alkali technology ranks second with respect to total gas flow being controlled by FGD systems now operating, under construction, or planned (see Table 2). The process offers the advantage of SO_2 removal by soluble

alkali salts (usually sodium salts), with the concomitant elimination of scaling and plugging problems in the absorber. The soluble scrubber effluent is then treated with lime (or limestone) to regenerate the scrubbing liquor and produce a solid product, calcium sulfite/sulfate, for disposal.

This paper will not discuss dual alkali applications in any depth, because of other papers being presented at this symposium on the Firestone, Pottstown and the General Motors, Parma installations. However, a brief summary of other dual alkali systems follows. Table 6 summarizes these plants.

Caterpillar Tractor Company, Illinois

Caterpillar has purchased dual alkali FGD systems from Zurn Industries and from FMC Environmental Equipment Division. The first system to begin operation was a Zurn dilute-mode system at the Joliet plant in September 1974. Since then the following Caterpillar plants have initiated FGD operations: Mossville (FMC concentrated-mode, October 1975), Morton (Zurn dilute-mode, January 1978), and East Peoria (FMC concentrated-mode, April 1978). The Mapleton plant was scheduled to begin operating an FMC system in January 1979.

All the Caterpillar heat plants fire a high sulfur (2.5-3.2 percent) Illinois coal. In general, recurrent problems with the systems in the Caterpillar plants have yet to be solved. Although all of these facilities rely on the FGD system for some particulate removal, the systems have yet to yield fully satis-

TABLE 6. INDUSTRIAL SITES USING DUAL ALKALI FGD TECHNOLOGY

Company	Location	Fuel	Flow rate scfm	Reagent	Vendor	% SO ₂ removal
Caterpillar Tractor Co.	E. Peoria, IL	Coal	210,000	Dual alkali (Concentrated)	FMC Env. Equipment	90
Caterpillar Tractor Co.	Joliet, IL	Coal	67,000	Dual alkali (Dilute)	Zurn Industries	90
Caterpillar Tractor Co.	Morton, IL	Coal	38,000	Dual alkali (Dilute)	Zurn Industries	90
Caterpillar Tractor Co.	Mossville, IL	Coal	140,000	Dual alkali (Concentrated)	FMC Env. Equipment	90+
Firestone Tire & Rubber Co.	Pottstown, PA	Coal	8,070	Dual alkali (Concentrated)	FMC Env. Equipment	90.5
General Motors Corp.	Pama, OH	Coal	128,400	Dual alkali (Dilute)	GM Environmental	90

factory performance in this area. An exception is the Morton installation where some in-house modifications have improved overall operations.

Caterpillar has not released cost data for these systems.

Other Systems

The remaining dual alkali FGD systems covered in the EPA Industrial Boiler FGD Survey are either under construction or planned.

As mentioned earlier, Chanslor Western Oil and Development Co. in Bakersfield, California has purchased a system from FMC; system startup is scheduled for the spring of 1979. This system is designed for 96 percent SO₂ removal (710 ppm at the inlet), and will control 133,400 acfm of flue gas at 550°F from eight steam generators totaling 310 million Btu/h.

Arco/Polymers, Inc. in Monaca, Pennsylvania is planning on a June 1980 startup for its FMC dual alkali system. The boiler plant at Monaca consists of three coal-fired units (firing 3.0 percent sulfur coal) which generate 900,000 lb/h of steam at 700 psig and 700°F. This system is designed for 90 percent SO₂ removal (1800 ppm at the inlet).

The U.S. Air Force has purchased an Airpol Industries dual alkali system for its Grissom Air Force Base near Bunker Hill, Indiana. This system is scheduled for startup in September 1979. It will control emissions from three coal-fired (0.7 to 4.6 percent sulfur) boilers.

Dupont, Inc. is planning a coal-fired (1.5 percent sulfur) installation at an as yet unspecified site in Georgia. The boiler plant will consist of three units generating a total of 1.2 million lb/h of steam. Regulations are expected to require 90 percent SO₂ removal. Startup is scheduled for 1987.

ALKALI WASTE STREAM FGD TECHNOLOGY

Pulp mills, textile mills, and beet sugar plants may use an onsite waste stream as the scrubbing liquor for SO₂ removal. This would hold true for any industrial operation where a high-pH waste stream is available in sufficient quantity. Industrial sites using alkali waste streams for SO₂ control are:

- ° American Thread, Marion, NC -- caustic waste stream; coal (1.0-1.5 percent sulfur); 18,500 scfm; vendor: W.W. Sly Manufacturing.
- ° Canton Textiles, Canton, GA -- caustic waste stream; coal (0.8 percent sulfur); 25,000 scfm; vendor: FMC.
- ° Georgia-Pacific Paper Co., Crossett, AR -- caustic waste stream; bark, coal, and oil (1.5-2.0 percent sulfur); 220,000 scfm; vendor: Airpol.
- ° Great Southern Paper Co., Cedar Springs, GA -- caustic waste stream; bark, coal, and oil (1.0-2.0 percent sulfur); 420,000 scfm; vendor: Airpol.
- ° Great Western Sugar, several locations -- ammoniacal waste stream; coal (0.6-2.0 percent sulfur); 462,000 scfm; in-house design.
- ° Minn-Dak Farmers' Co-op, Wahpeton, ND -- ammoniacal waste stream; lignite (1.0 percent sulfur); 164,000 scfm; same design as Great Western Sugar.
- ° Nekoosa Papers, Inc., Ashdown, AR -- caustic waste stream; coal (1.0-1.5 percent sulfur); 211,000 scfm; vendor: Airpol.

- ° St. Regis Paper Co., Cantonment, FL -- caustic waste stream; bark, oil, and gas (less than 1.0 percent sulfur); 115,000 scfm; vendor: Airpol.

OTHER FGD TECHNOLOGIES

Table 7 summarizes plants using other FGD technologies.

Lime/Limestone--

This is not a predominant technology in industrial FGD applications. The two operating systems are at Rickenbacker Air Force Base near Columbus, Ohio, and at Armco near Middletown, Ohio.

Bureau of Mines Citrate--

St. Joe Zinc Co., in Monaca, Pennsylvania will soon be initiating the startup phase of Citrate FGD system operations. This FGD system, discussed in another paper at this symposium, will be the first full-scale (60-MW) demonstration for this promising regenerable process.

SULF-X--

This regenerable acid mine waste process has been piloted on a 1-MW slipstream at Fort Benjamin Harrison. A larger scale demonstration (10,000 scfm or about 5 MW equivalent) will begin this year at the Western Correctional Institute in Pittsburgh. Startup of the absorber is scheduled for May 1979 and sulfur production will begin in late fall 1979. This system has demonstrated the potential for simultaneous SO_2 and NO_x removal. The process is offered by Pittsburgh Environmental and Energy Systems.

TABLE 7. INDUSTRIAL SITES USING OTHER FGD TECHNOLOGIES

Company	Location	Fuel	Flow rate, scfm	Reagent	Vendor	% SO ₂ removal
Armco Steel	Middletown, OH	Coal	84,000	Lime	Koch Engineering	--
Rickenbacker Air Force Base	Columbus, OH	Coal	55,000	Limestone	Research-Cottrell/Bahco	90+
St. Joe Zinc Co. ^a	Monaca, PA	Coal	142,000	Citrate	Bureau of Mines	90+
Western Correctional Institute ^b	Pittsburgh, PA	Coal	10,000	Ferric sulfide	Pittsburgh Env. and Energy	90

^a Under construction.^b Planned, contract awarded.

CAPITAL AND ANNUAL COSTS

Table 8 summarizes available data on capital and annual costs for nonregenerable sodium and alkaline waste stream FGD systems operating on industrial boilers. Sufficient data was not available for the other technologies to report meaningful averages.

The cost figures reported by the plants were adjusted by PEDCo for inflation to 1978 dollars. In some cases adjustments were made for equipment, installation, instrumentation, and engineering. These costs are not to be taken as representing the product of a complete cost analysis. They are presented to provide a guideline for the costs associated with these two representative FGD processes. The capital costs are expressed in terms of dollars/ scfm because the amount of flue gas to be treated is the most critical design parameter in sizing an FGD system. The operating costs, expressed as ¢/10⁶ Btu, are based on 340 days operation at 100 percent load. Realistic cost figures will be higher since, in actual operation, a yearly capacity factor could be as low as 20 to 30 percent.

TABLE 8. ADJUSTED CAPITAL AND ANNUAL COSTS

Process	\$/scfm (number of plants)	¢/10 ⁶ Btu (number of plants)
Nonregenerable sodium	15.50 (13)	15.49 (5)
Alkaline waste streams	9.68 (4)	2.42 (2)

SUMMARY/CONCLUSIONS

Existing FGD applications in the industrial sector have generally shown that high reliabilities of system operation can be attained along with SO₂ removal efficiencies as high as 98 percent. There has been a steady increase in the application of industrial FGD, with an average of about seven new systems starting up each year since 1972. Flue gas capacities of known operational and planned FGD systems totals over 9 million scfm (4600 MW equivalent). Regenerable as well as the more commonly used nonregenerable technologies are being developed by a number of FGD system vendors with varying degrees of success. The new more stringent regulations for SO₂ emissions may be expected to increase interest in these technologies, and to stimulate performance of future FGD systems.

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1. Melia. M., et al., EPA Utility FGD Survey: October-November 1978, EPA-600/7-79-022b, PEDCo Environmental, Inc., Cincinnati, Ohio, January 1979.
2. Tuttle, J.D., et al., EPA Industrial Boiler FGD Survey: Fourth Quarter 1978, EPA-600/7-79-067a, PEDCo Environmental, Inc., Cincinnati, Ohio, February 1979.
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SUPPLEMENTARY INFORMATION

The British system of measurement is used in this report. Some of the conversion factors between the British and metric systems are shown below:

<u>British Unit</u>	<u>Metric Unit</u>
1 t (short ton)	0.91 metric ton
1 lb (pound)	0.45 kilogram
1 gal. (gallon)	3.79 liter
1 scfm	1.58 normal cubic meter/hour
1 gal./1000 acf	0.134 liters/actual cubic meter
1 Btu	1.05 x 10 ³ joules

In addition, certain engineering terms used in this paper are defined below:

acf	Actual cubic feet, unit of gas volume measured at its actual temperature and pressure
acfm	Actual cubic feet per minute, unit of gas flow rate measured at its actual temperature and pressure
Availability	Hours the FGD system was available (whether operated or not) divided by hours in period, expressed as percentage
Reliability	Hours the FGD system was operated divided by hours it was called upon to operate, expressed as percentage

ENVIRONMENTAL ASSESSMENT OF THE DUAL ALKALI FGD SYSTEM
APPLIED TO AN INDUSTRIAL BOILER FIRING COAL AND OIL

Wm. H. Fischer
Gilbert Associates
Reading, Pa.

Wade H. Ponder
U.S. Environmental
Protection Agency
Industrial Environmental
Research Laboratory
Research Triangle Park, N.C.

Roman Zaharchuk
Firestone Tire and Rubber Co.
Pottstown, Pa.

ABSTRACT

This paper summarizes the results of a comparative multimedia assessment of a dual alkali flue gas desulfurization system on a coal/oil industrial boiler to determine relative environmental, energy, economic, and societal impacts. Comprehensive sampling and analyses of multimedia emissions from the boiler and its control equipment were conducted to identify criteria pollutants and other species. The results indicate that: (1) while the quantity of uncontrolled particulate matter from oil-firing is considerably less than from coal-firing, the oil-fired particles are generally smaller and the concentration of particles in the treated flue gas from oil-firing is approximately the same as from coal-firing; (2) uncontrolled emissions of NO_x and CO during coal-firing are about triple those during oil-firing; (3) while sulfate emissions from the boiler during coal-firing are about triple those during oil-firing, sulfate emissions after the control equipment are essentially identical; (4) emissions of cadmium, calcium, magnesium, nickel and vanadium are higher during oil firing; (5) oil-firing may produce cadmium concentrations in vegetation approaching levels injurious to humans; coal-firing may produce molybdenum levels injurious to cattle.

INTRODUCTION

Objective

Conventional methods of converting fossil fuels to usable forms of energy have impacts on the air, land and water, i.e., "multimedia impacts." These impacts are not separate and distinct; rather, they are interrelated and involve delicate balances and trade-offs.

The Environmental Protection Agency (EPA), with primary responsibility for controlling adverse environmental impacts of pollutant emissions, has been active since its inception in determining the identities and quantities of potential pollutants released to the environment when fossil fuels are burned. Information from EPA R&D efforts is being used for three principal purposes: to assess the health and environmental impacts caused by the release of combustion pollutants to the environment; to define the needs for technology to control the release of these pollutants; and to develop standards to limit emissions.

CCEA Program

In response to the need for a comprehensive environmental assessment of conventional combustion systems, EPA's Industrial Environmental Research Laboratory at Research Triangle Park (EPA/IERL-RTP), North Carolina, has established a unified Conventional Combustion Environmental Assessment (CCEA) program. It is a major new program aimed at the comprehensive assessment of environmental, economic, and energy impacts of multimedia pollutant emissions from stationary industrial, utility, residential, and commercial conventional combustion processes. The primary objective of the CCEA program is to identify and evaluate information from all relevant sources in order to: determine the extent to which this information can be utilized to assess the total environmental, economic, and energy impacts of conventional combustion processes; identify and acquire additional information needed for such assessment; define the requirements for modifications or additional development of control technology; and define the requirements for modified or new standards to regulate pollutant emissions.

The CCEA program will coordinate and integrate ongoing and future studies into a unified environmental assessment structure and serve as a centralized base of information on the environmental impacts of conventional combustion processes. Coordination and information exchange between CCEA-related studies should minimize duplication and maximize the return from available resources.

The environmental assessment (EA) methodology employed in the CCEA program fundamentally consists of three iterative steps (Figure 1):

1. Characterization of the combustion process (including any associated pollution control devices) and its effluents.

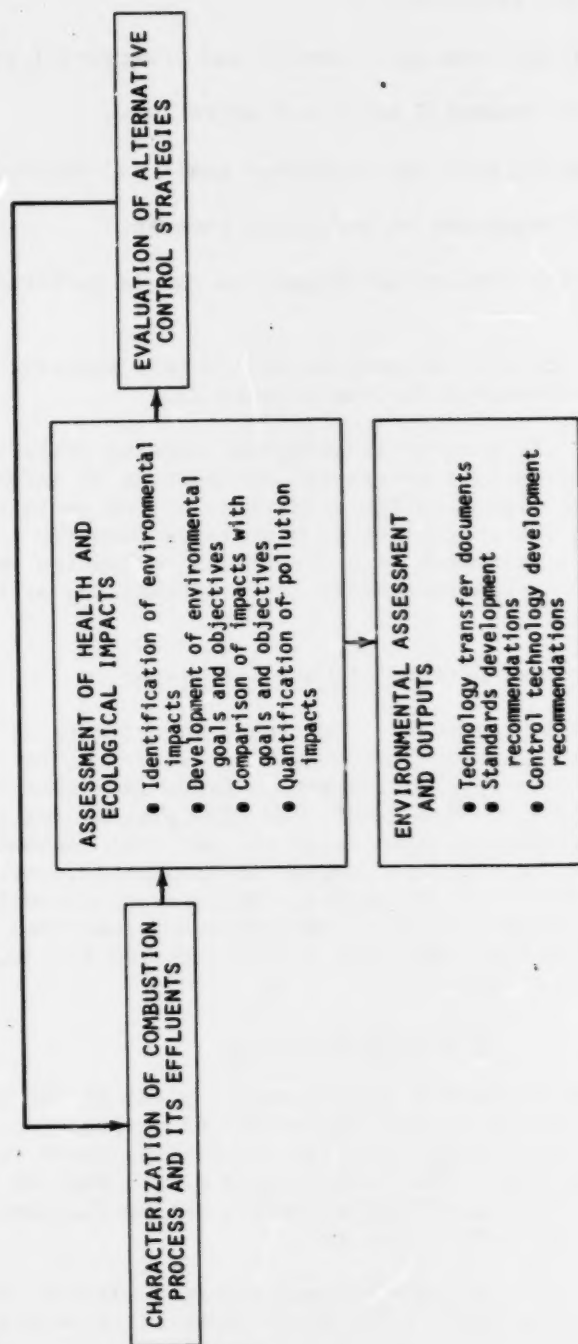


FIGURE 1
GENERALIZED ENVIRONMENTAL ASSESSMENT METHODOLOGY

2. Assessment of the health and ecological impacts of the combustion process and its effluents on the environment:

Identification of environmental (health and ecological) impacts.

Development of environmental goals and objectives.

Comparison of impacts with environmental goals and objectives.

Assessment of the magnitude of pollution impacts.

3. Evaluation of alternative control strategies to reduce pollution impacts to acceptable levels.

The EA procedure used in the CCEA program and in this specific study is shown in the generalized methodology diagram (Figure 2).

It is the goal of the CCEA program to integrate ongoing projects and recommend new efforts to address all practical combinations of information. It is expected that EPA/IERL-RTP, with the assistance of contractors with experience and expertise in the various areas associated with the comprehensive environmental assessment of conventional combustion processes, will implement and expand the CCEA program as needs dictate and as resources permit.

Multimedia Effects of Coal Conversion

A major goal of the CCEA program is to evaluate the effects of implementation of the National Energy Plan, which calls for the increased use of coal to meet the Nation's energy requirements. Since fuel switching from oil to coal is an important facet of the NEP, the CCEA program initiated a study to evaluate the environmental effects of oil and coal combustion in a controlled industrial boiler in order to compare environmental, energy, and societal impacts of firing coal vs. firing oil. In order to conduct the comparative assessment, it was necessary to fully characterize feed streams, emissions, and effluents from the industrial boiler selected for study and all associated pollution control equipment.

Plant Description

The site chosen was the Pottstown, Pennsylvania, plant of the Firestone Tire and Rubber Company, with Firestone's agreement and cooperation. Boiler No. 4, one of four boilers which supply process and heating steam to the plant, was used in the assessment. The boiler burns either coal or oil and has a pilot FMC dual alkali flue gas desulfurization system designed to treat approximately one-third of the boiler flue gas.

Boiler No. 4 is a dry, bottom, once-through integral furnace, Babcock and Wilcox (Type FH-18) unit. (See Table 1 for boiler specification data and Figure 3 for a schematic of the boiler and associated equipment.) When it was installed in 1958, the boiler was designed as a coal-fired unit but was converted to fire either coal or oil in 1967. The changeover from one fuel to the other can be accomplished in less than 30 minutes.

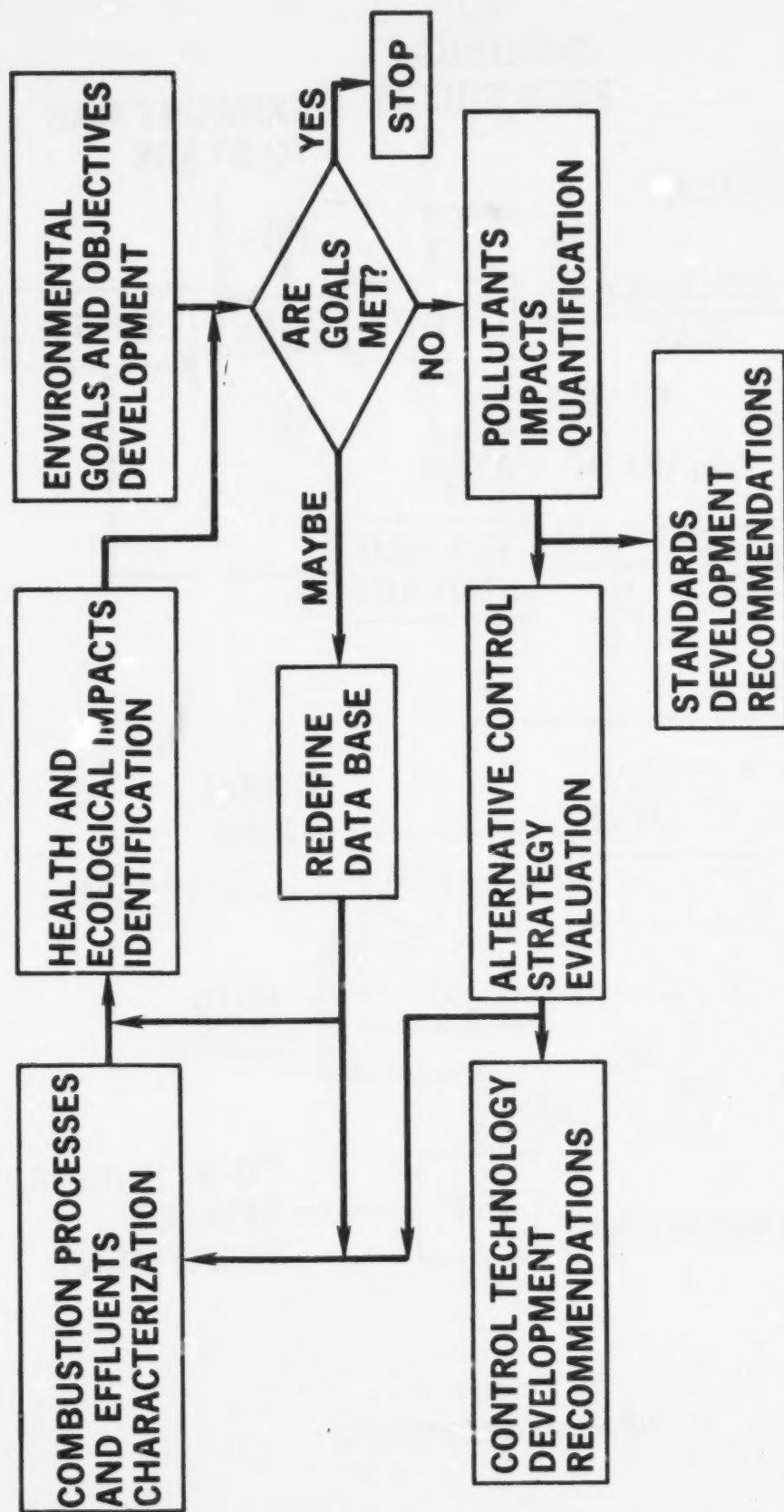


FIGURE 2
GENERAL METHODOLOGY

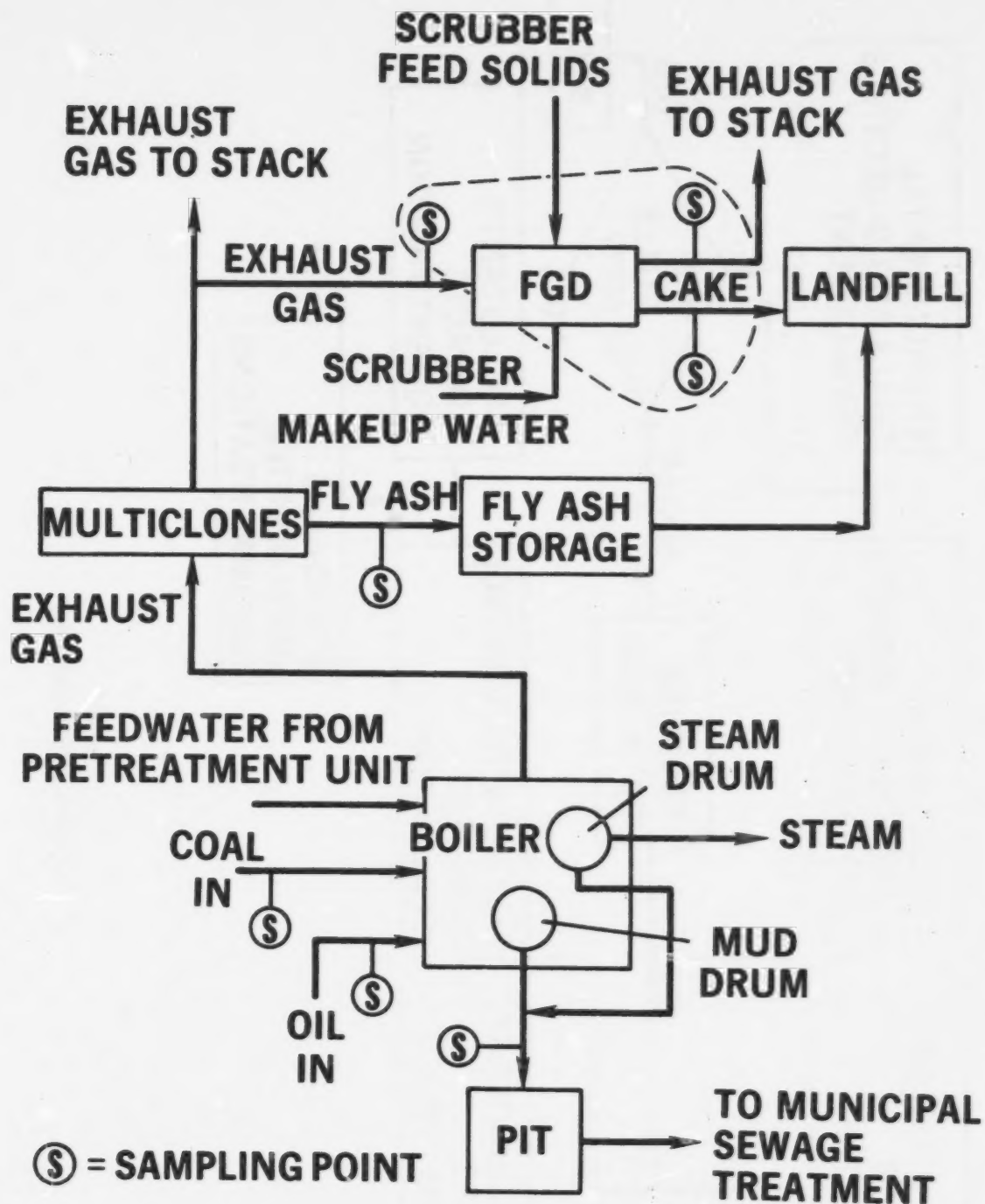


FIGURE 3
BOILER SYSTEM SCHEMATIC

For test purposes, Firestone agreed to fire one fuel and then the other as long as required to conduct the appropriate sampling.

TABLE 1 BOILER NUMBER 4 DESIGN DATA

Boiler Type:	Oil/Pulverized coal; face fired; integral furnace; dry bottom
Manufacturer:	Babcock and Wilcox, Type FH-18
Type of Burner:	Circular conical
Number of Burners:	3
Burner Arrangement:	Triangular, one face
Air Preheater:	Yes
Fuel:	No. 6 fuel oil; High volatile Pennsylvania bituminous coal; Class II, Group 2, of ASTM D388; from Island Creek Coal Co.
Design Steam Rate:	45,000 kg/hr (100,000 lb/hr); 1.4 MPa (190 psi); at approximately 193°C (380°F)
Use:	Process steam

The two fuels are usually not burned simultaneously except when converting from oil to coal firing. The coal is ignited by continuing oil firing until a stable coal flame is obtained. Oil is fired simultaneously with coal to maintain acceptable steam generation rates when coal with a low heat content is burned. Fuel analyses are given in Table 2.

Control Devices

The flue gases are treated by an air pollution system which consists of multiclone units and a pilot FGD unit. The multiclones are the primary particle control device. All of the flue gas passes through the multiclones after which the stream is split: two-thirds of the flue gas is ducted to the stack; and the other one-third is ducted to the pilot FGD system which removes SO₂ and additional particles. There are no NO_x controls on the system.

TABLE 2 SUMMARY OF AVERAGE ULTIMATE COAL AND OIL ANALYSES

<u>Species</u>	<u>Coal</u>	<u>Weight %</u>		<u>σ^a</u>
		<u>σ^a</u>	<u>Oil</u>	
Moisture	7.15	0.86	--	--
Carbon	72.10	1.07	86.28	0.39
Hydrogen	4.28	0.06	10.92	0.03
Nitrogen	0.92	0.07	0.36	0.06
Chlorine	0.12	0.02	--	--
Sulfur	1.64	0.23	1.96	0.08
Ash	9.90	0.85	0.02	0.004
Oxygen	3.89	0.23	0.46	0.40
kJ/kg (Btu/lb)	29,485 (12,686)	459	40,741 (17,528)	--

σ^a One standard deviation.

The collection efficiency of the multiclone varies as a function of the particle size distribution and grain loading. Typically, multiclones remove 90 percent of those particles with diameters of 10 μ m and greater, and 50 to 80 percent of those particles with diameters of 3 μ m and greater. The collection efficiency of multiclones drops off rapidly for particles less than 3 μ m diameter.

The flue gas desulfurization (FGD) system was designed and manufactured by FMC Corporation. The FGD system is a pilot unit designed to handle 280 acm/min (10,000 acfm) of flue gas, which is approximately one-third of the volume of the flue gas from the boiler. The pilot plant was placed on-line in January 1975. Figure 4 is the basic flow diagram of the FMC FGD system, as applied at this site.

The flue gas (stream 1) is withdrawn downstream of the boiler on the exit side of the multiclone dust collectors. Fly-ash loading at the scrubber inlet is substantially higher during coal-firing than during oil-firing. To accommodate the wide variation in fly-ash loading, the FGD system was designed to operate with or without fly-ash, and can be operated on either fuel.

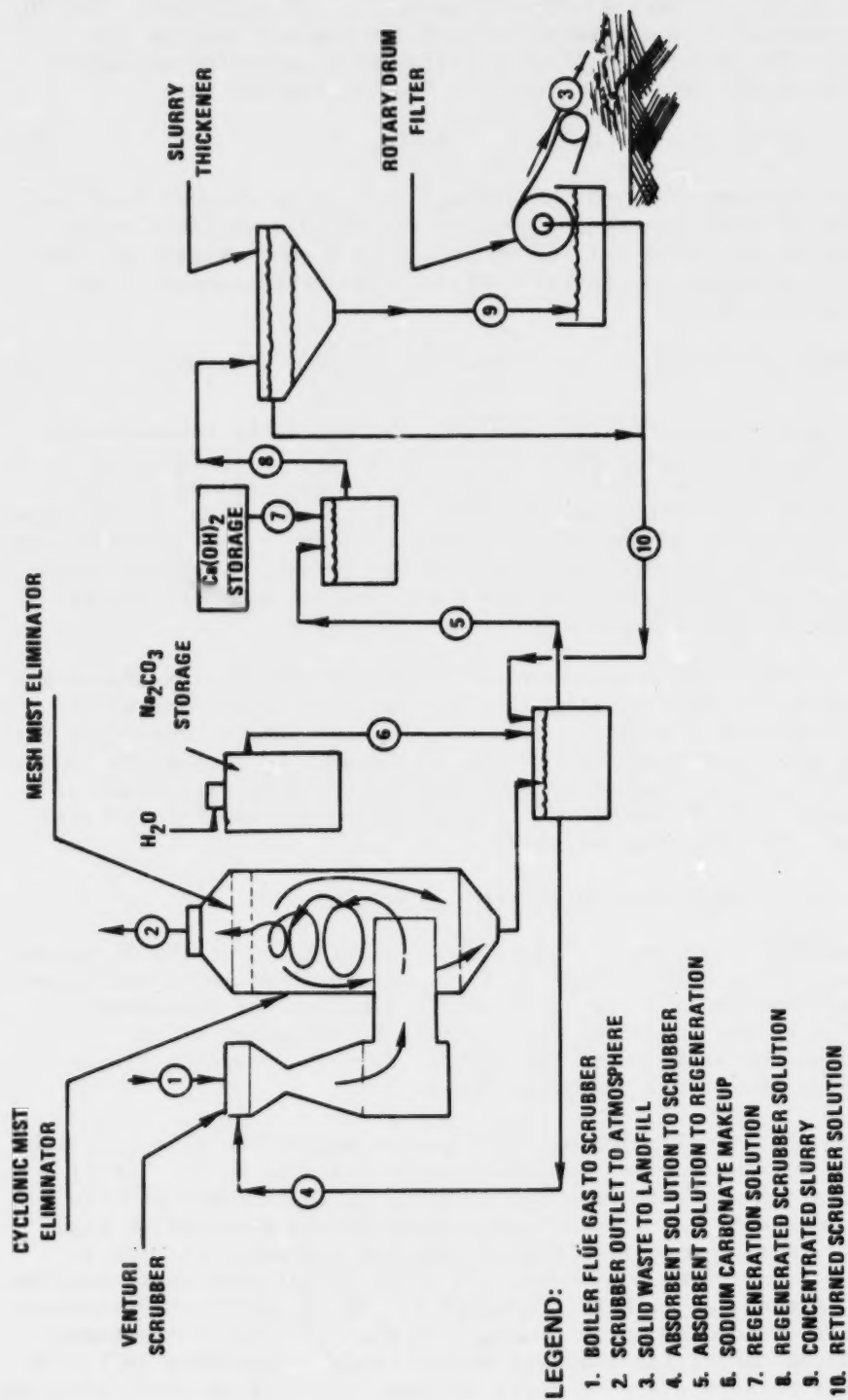
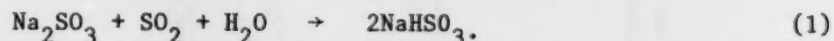
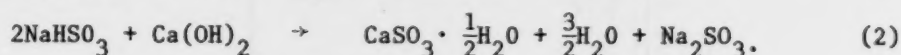


FIGURE 4
FMC UNIT AT THE INDUSTRIAL FACILITY

Upon entering the FGD unit, the flue gases are contacted with a slightly acidic scrubbing solution (stream 4) which removes SO_2 and particles. The SO_2 and particles are removed at the scrubber throat and carried away in the scrubbing solution. The process utilizes a sodium sulfite/sodium bisulfite solution as the absorbent. The basic reaction for SO_2 removal is:



A bleed stream (stream 5) of the scrubbing solution is removed from the system at a rate which keeps the pH of the solution in an acceptable range. The bleed stream is reacted with calcium hydroxide in a short retention time, agitated vessel to regenerate the sodium sulfite. The basic chemistry of sodium sulfite regeneration is:



The slurry of precipitated sulfur compounds (stream 8) is concentrated and pumped to a rotary drum filter where the essentially clear liquid is separated from the solid waste products. The clear liquid (stream 10) is returned to the system for further utilization. The solid wastes, in the form of filter cake containing 40 weight percent water (stream 3), are removed from the rotary drum filter and conveyed to a storage bin to await transportation to the dump site. Mainly due to the heavier particle loading, more filter cake is produced during coal firing than during oil firing.

The on-site landfill, which is the final disposal facility for flyash and scrubber cake generated at the facility, has several test wells from which samples are collected every 3 months and sent to an independent laboratory for analysis. Monthly tests are conducted by plant personnel to monitor Na^+ and specific conductivity. With permission from the Pennsylvania Department of Environmental Resources, this site is being used as an experimental disposal area for the filter cake from the FMC unit.

Test Description and Conditions

Multimedia emission tests were conducted on Boiler No. 4 of the Firestone Plant from September 27 through October 8, 1977. Gaseous and solid emissions were sampled during coal and oil firing to obtain data for the assessment. Flue gas was sampled before and after the scrubber to determine which pollutants are removed or modified by the control device. Sampling points used are indicated on the process diagram, Figure 3.

Emissions were characterized using EPA's phased approach. This approach utilizes two levels of sampling and analysis (Level 1 and Level 2). Level 1 screening procedures are accurate within a factor of 2 to 3. They provide preliminary assessment data and identify problem areas and information gaps. Based on these data, a site specific Level 2 sampling and analysis plan is developed. Level 2 provides more accurate and detailed information to confirm and expand on the information gathered in Level 1. The methods and procedures used for Level 1 are documented in the manual, "Combustion Source Assessment Methods and Procedures Manual for Sampling and Analysis", September 1977, (in press). The Level 2 methods and procedures include "state-of-the-art" techniques required for this particular site.

Normally all Level 1 samples are analyzed and evaluated before moving to Level 2. Because of the program time constraints, the Level 1 and Level 2 samples were obtained during the same test period; however, analysis of the samples did proceed in a phased manner except where sample degradation was of concern. In that case, Level 2 analysis was performed on the sample prior to Level 1 completion.

Gaseous Effluents

The boiler flue gas was sampled at the inlet and the outlet of the pilot flue gas desulfurization unit. Integrated bag samples were taken at both points during each test. On-site analyses of CO_2 , O_2 , N_2 and $\text{C}_1 - \text{C}_6$ organics were conducted on the bag samples. Continuous monitors were used to analyze CO , NO , NO_2 , NO_x , SO_2 and total hydrocarbons (as CH_4). Figure 5 is a schematic of the continuous monitor setup. A Thermal Electron Corporation (TECO) gas conditioner was used to remove condensate and particulate matter from the gas sample. Gaseous streams were isokinetically sampled at each location during all tests using four different sampling trains.

The Source Assessment Sampling System (SASS) was used to collect Level 1 gaseous and particulate emission samples at the scrubber inlet and outlet. The SASS train is illustrated in Figure 6. The train consists of a heated probe, three cyclones and a filter in a heated oven. The cyclones were used only during the coal inlet tests. During the other tests, the particle loadings were too low for the cyclones to work effectively. The remainder of the system consists of a gas conditioning system, an XAD-2 polymer absorbent trap and a series of impingers. The polymer traps gaseous organics and some inorganics and the impingers collect the remaining inorganics. All sample contact surfaces are Type 316 stainless steel, Teflon, or glass. The train was run for 6 to 8 hours so that a minimum of 30 cubic meters of gas was collected.

Previous sampling and analysis experience had indicated that SASS train materials may contaminate certain organic and inorganic samples. The contamination is of concern only when the pollutant is present at a concentration that is near the detection limit of the Level 2 methods. To avoid that possibility, all glass sampling trains were used to collect Level 2 samples. Method 5 sampling trains were modified as shown in Figure 7 for organics and Figure 8 for inorganics. Both trains sampled approximately 10 cubic meters of flue gas during a 6 to 8 hour test run.

A controlled condensate train (Goksoyr-Ross), shown in Figure 9, was used at each location to obtain samples for SO_2 , SO_3 (as H_2SO_4), particulate sulfate, HCl , and HF .

During Level 2 test runs, Andersen cascade impactors were used to obtain particle samples by particle size fraction. A pre-separating $10\mu\text{m}$ cyclone was used up-stream of the impactor on the inlet side.

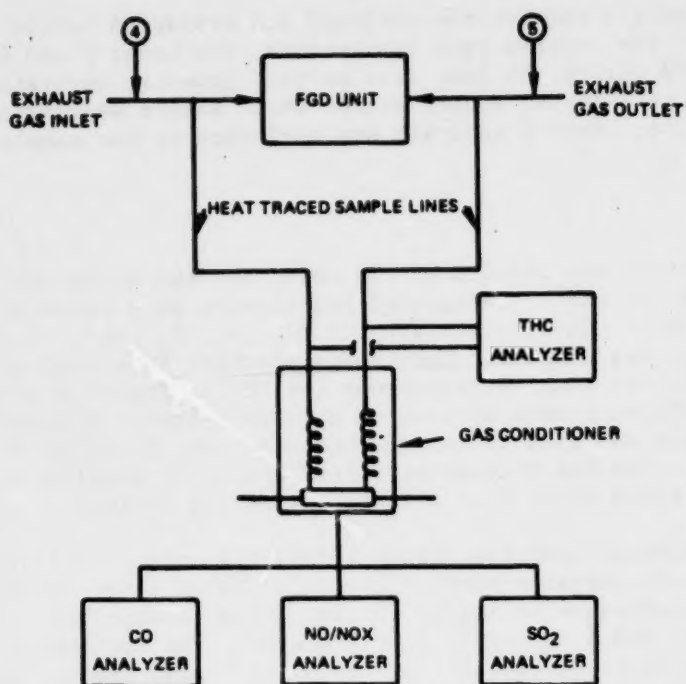


FIGURE 5
FLUE GAS CONTINUOUS MONITOR SETUP

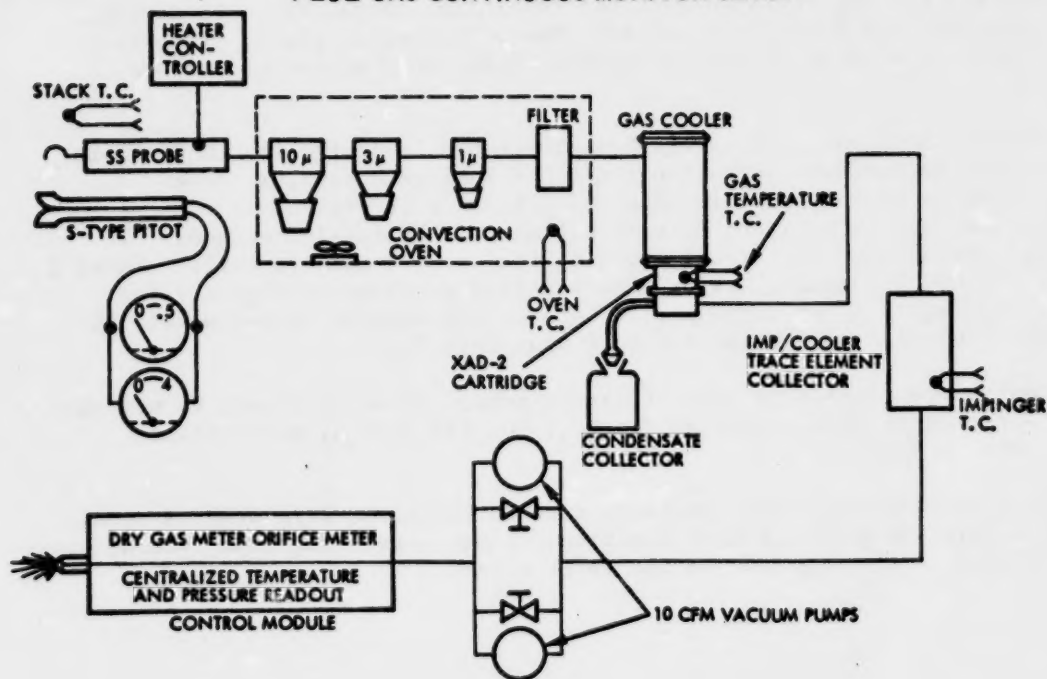


FIGURE 6
SOURCE ASSESSMENT SAMPLING SYSTEM (SASS) SCHEMATIC

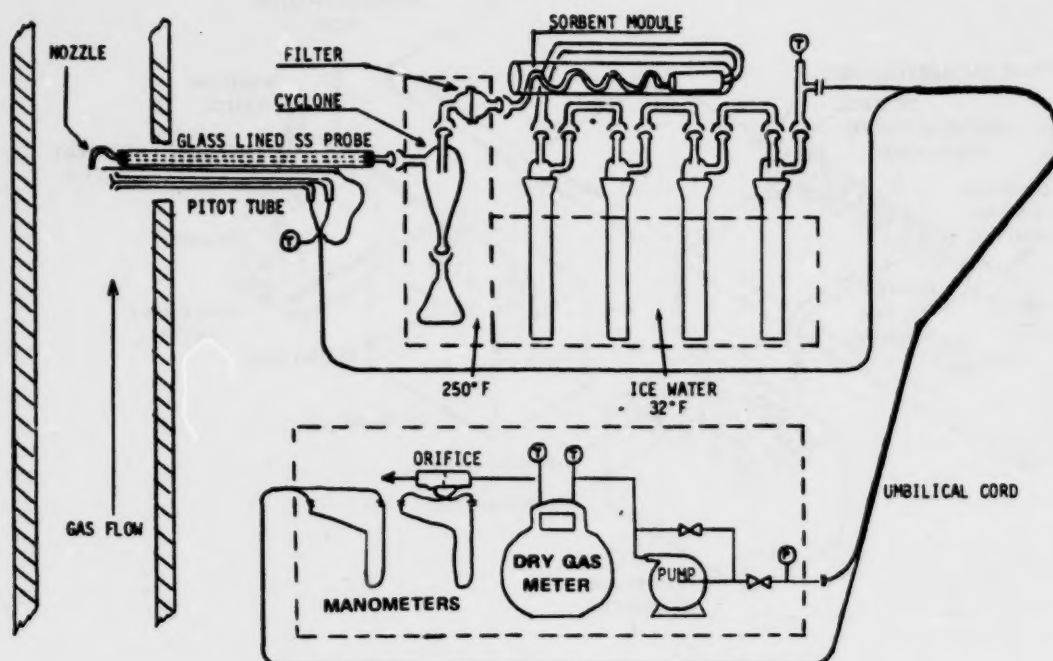


FIGURE 7
ORGANIC SAMPLING TRAIN

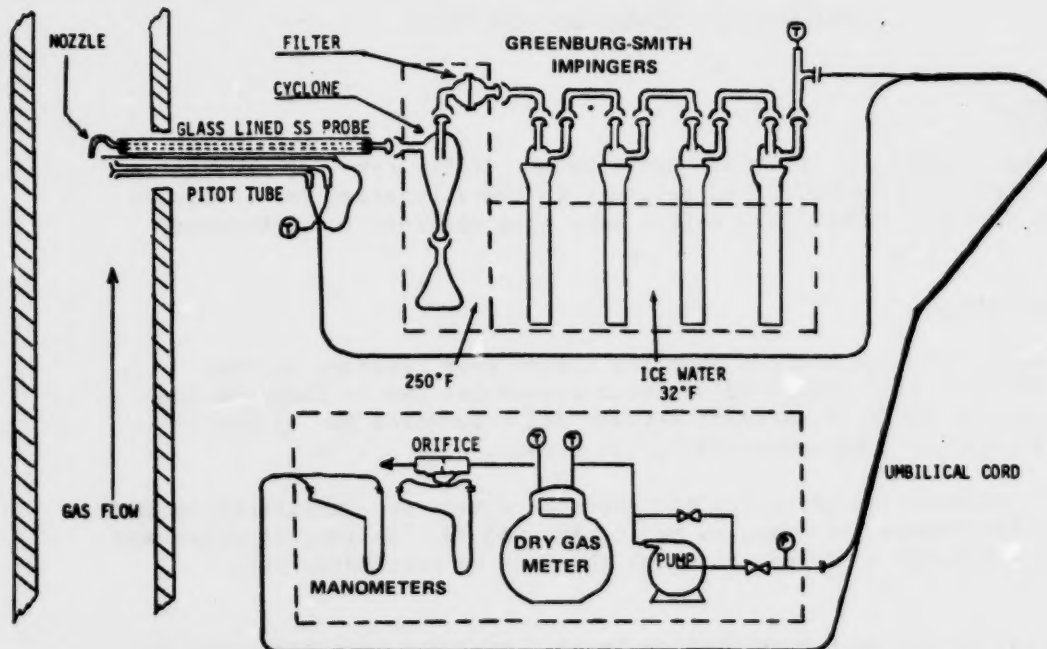


FIGURE 8
INORGANIC SAMPLING TRAIN

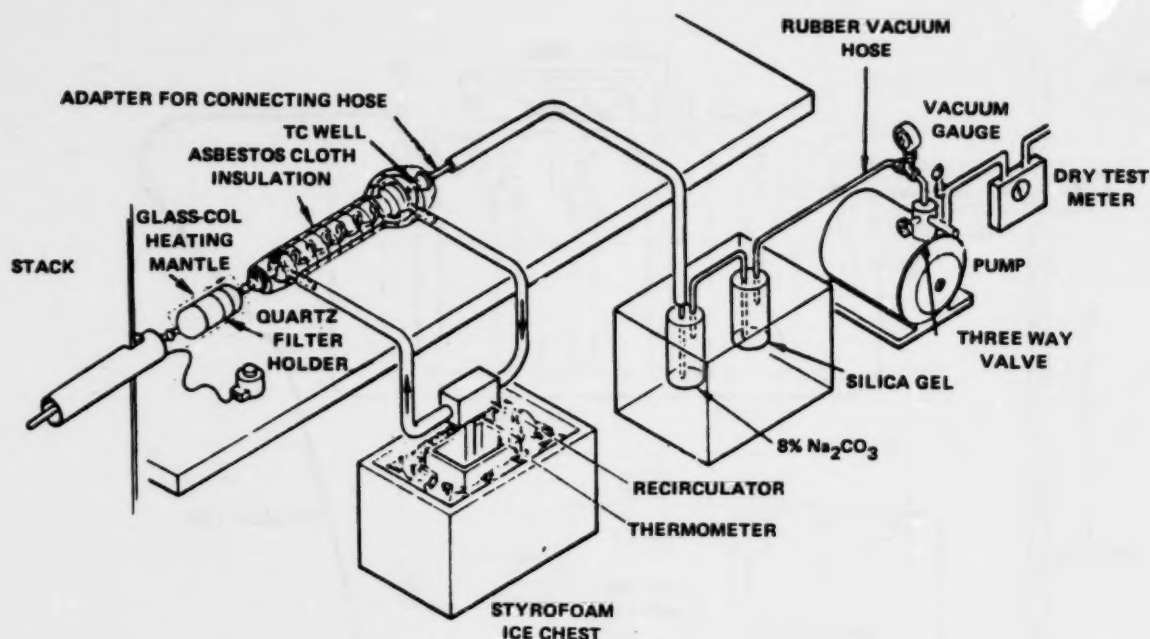


FIGURE 9
CONTROLLED CONDENSATION TRAIN

Solid Effluents

Composite samples of the flyash and scrubber filter cake were collected according to Level 1 procedures and returned to the laboratory for analysis. Grab samples of the scrubber feed solids were also obtained for laboratory analyses.

Laboratory Analyses

The samples from the various sampling trains were returned to the laboratory for analysis. Detailed analysis procedures can be found in the manual "Combustion Source Assessment Methods and Procedures Manual for Sampling and Analysis", September 1977, (in press).

Level 1 analyses for particles and gases were made for inorganics by SSMS and for selected anions and organics by LC, IR, and MS. Solids, slurries and liquids were similarly analyzed, although the work-up procedures were different.

More detailed and more quantitative Level 2 analyses were performed to identify and quantitate specific compounds indicated by the Level 1 analyses.

TABLE 3 SUMMARY OF TEST CONDITIONS

Test No.	Steam Production Rate		% of Maximum Boiler Load	Nominal Fuel Feed Rate,		% O ₂ at Scrubber Inlet ^a	Estimated % Excess Air ^b to Furnace
	kg steam/hr	lb steam/hr		kg/hr	gal/hr		
COAL FIRING							
200	39,700	87,500	87.5	3629		7.8	20
201-1	44,200	97,500	97.5	3629		8.2	20
201-2	43,100	95,000	95.0	3629		8.4	20
201-3	34,000	75,000	75.0	3175		8.3	20
201-4	40,800	90,000	90.0	3629		6.7	20
OIL FIRING							
202-1	45,000	100,000	100		900	5.8	21
202-2	45,400	100,000	100		900	6.3	21
202-3	44,200	97,500	97.5		880	6.1	21
202-4	42,200	93,000	93.0		805	4.0	21
203	31,800	70,000	70.0		600	Not Measured	21

^aDue to air leaks in ducting upstream of the scrubber inlet, tabulated O₂ values are not representative of combustion zone O₂ concentrations, which normally range from 3 to 4% for this unit.

^b% excess air is estimated to be $100 \times \frac{O_2 - CO/2}{0.264 N_2 - (O_2/2)}$, where O₂ was assumed to be

3.5% and other species concentrations were computed from fuel analysis.

Test Conditions

Ten tests were performed with the industrial boiler, five each with coal and oil. Unit loadings ranged from 31,800 to 45,400 kg steam per hour (70,000 to 100,000 lb per hour), which corresponds to between 70 and 100 percent of full load operation. Specific test conditions are summarized in Table 3.

Test data relating to scrubber throughput and loading and total flue gas generation rates are presented in Table 4. The scrubber is a pilot unit which does not process the entire flue gas output of the furnace. From 11 to 32 percent of the total flue gas was processed through the scrubber during the tests. Typical inlet and outlet gas temperatures for the scrubber unit were 149°C and 52°C (300°F and 125°F). Only 51 to 57 percent of design loading, rather than full loading, was maintained during coal-fired testing because failure of the multiclone particle removal system upstream of the scrubber resulted in high solids loading at the scrubber and unacceptably high scrubber filter-cake production rates. During oil-fired testing, 88 to 100 percent of full design flows were maintained.

Analytical results were used to estimate total boiler emission on the basis of treatment of 100 percent of the flue gas from the boiler. That is, it was assumed that additional scrubber modules could be added in parallel to the system such that the total flue gas output would be processed with a mean scrubbing efficiency equal to that of the pilot scrubber.
All stack emissions data are based on this assumption.

TABLE 4 FRACTION OF FLUE GAS PROCESSED BY SCRUBBER

	Flow Rate at Scrubber Inlet dscm/min ^a	% of Design Load	Total Flue Gas Flow Rate ^a dscm/min	Fraction of Total Flue Gas Processed by Scrubber
COAL FIRING				
Average	96	54	741	0.13
OIL FIRING				
Average	180	102	737	0.25

^a dry standard cubic meters per minute

MULTIMEDIA EMISSION RATES

Gaseous Emissions

Particulate Matter

Particle concentration during coal firing at the inlet to the scrubber averaged 2951 ng/J (6.86 lb/10⁶ Btu). Since the multiclone unit, upstream of the scrubber, failed during the test period and thus removed little or no particulate material, this may be taken to be representative of uncontrolled emissions. The outlet concentration after scrubbing was 18.6 ng/J (0.04 lb/10⁶ Btu), which corresponds to 99.4 percent particle removal efficiency.

The uncontrolled particle emission rate during oil firing was 113 ng/J (0.26 lb/10⁶ Btu). After scrubbing it was 17.6 ng/J (0.04 lb/10⁶ Btu) at the outlet, for a removal efficiency of 84.4 percent. These data are summarized in Table 5.

Particle emissions after scrubbing are well below the existing NSPS of 43 ng/J (0.10 lb/10⁶ Btu) but slightly higher than the proposed limitation of 12 ng/J (0.03 lb/10⁶ Btu).

TABLE 5 PARTICLE CONCENTRATION
ng/J (lb/10⁶ Btu)

	<u>Coal</u>	<u>Oil</u>
Scrubber Inlet	2951 (6.86)	113 (0.26)
Scrubber Outlet	18.6 (0.04)	17.6 (0.04)
NSPS	43 (0.10)	
Proposed Limit	12 (0.03)	

Size Distribution

The particle size distribution at the scrubber inlet during coal firing was determined with a polarizing light microscope on a filter catch, due to the high loading. These number percent results were converted to aerodynamic diameter and weight percent by assuming that all particles had the same density. This is a reasonable assumption because the major components of the particle generated from coal combustion, aluminosilicates and iron oxides, are known to partition equally among small and large sizes. With the constant density assumption, the weight distribution in each size range would be proportional to the product of the number distribution and the particle volume representing the size range. The particle volume was calculated based on the geometric mean diameter for the size range.

Size distributions of particles from coal firing at the scrubber outlet and of particles from oil firing at both scrubber inlet and outlet were determined with an Andersen cascade impactor. The impactor calibration gives aerodynamic diameter and weight distribution directly.

Tables 6 and 7 show a significant change in the particle size distribution before and after scrubbing. Large particles are removed with greater efficiency than are small particles. Table 7 indicates that the mass emission rate of the smallest particles increases after scrubbing, suggesting that fine particles are generated within the scrubber.

TABLE 6 SCRUBBER INLET AND OUTLET PARTICLE SIZE DISTRIBUTION (Weight %)

Aerodynamic Diameter Size Range (μm)	Coal		Oil	
	Inlet	Outlet	Inlet	Outlet
<1	0.0017	62	20	83
1-3	0.041	30	1	12
3-10	2.24	7	74	5
>10	97.7	1	5	0

TABLE 7 PARTICLE EMISSION RATES AND REMOVAL EFFICIENCY

Aerodynamic Size Range (μm)	Coal			Oil		
	Inlet	-kg/hr- Outlet	% Removal Efficiency	Inlet	-kg/hr- Outlet	% Removal Efficiency
<1	0.0055	1.30	<0	4.48	2.27	49.2
1-3	0.13	0.63	<0	0.22	0.33	<0
3-10	7.3	0.15	97.9	16.6	0.14	97.4
>10	316.5	0.021	>99.9	1.12	0.00	100
Total	324.0	2.10	99.3	22.4	2.74	87.8

Sulfur Compounds

The average SO_2 emission rate from coal firing ahead of the scrubber was 1112 ng/J ($2.59 \text{ lb}/10^6 \text{ Btu}$) and after scrubbing was 36.3 ng/J ($0.08 \text{ lb}/10^6$), for a mean scrubber efficiency of 96.7 percent.

During oil firing, the SO_2 emission rates were 993 ng/J ($2.3 \text{ lb}/10^6$) ahead of the scrubber and 26.8 ng/J ($0.06 \text{ lb}/10^6 \text{ Btu}$) after the scrubber, for a mean scrubber efficiency of 97 percent.

In both cases, the SO_2 emissions after the FGD system were substantially below existing and proposed NSPS emission limit, see Table 8.

TABLE 8 SULFUR DIOXIDE EMISSIONS
ng/J ($\text{lb}/10^6 \text{ Btu}$)

	<u>Coal</u>	<u>Oil</u>
Scrubber Inlet	1112 (2.59)	993 (2.31)
Scrubber Outlet	36.3 (0.08)	26.8 (0.06)
NSPS	520 (1.2)	344 (0.80)
Efficiency, %	96.7	97

The scrubber removed 95-97 percent of the SO_2 during coal firing and 97-98 percent during oil firing. Only 32-33 percent of the SO_3 was removed during coal firing and 28-29 percent during oil firing. This relatively poor removal efficiency for SO_3 is an indication that SO_3 is either present as very fine aerosols in the scrubber inlet gas or is converted to very fine aerosols as the flue gas is rapidly cooled in the scrubber.

The removal rate for $\text{SO}_4^{=}$ was 88 percent during coal firing and 60 percent during oil firing, indicating that most of the $\text{SO}_4^{=}$ in the scrubber inlet is associated with larger particulates. However, combustion generated sulfates may not be simply passing through the scrubber. Because of the possibility that the $\text{SO}_4^{=}$ species from coal combustion may be changed by the scrubbing process, an analysis effort to determine the $\text{SO}_4^{=}$ species was initiated. Both the Fourier Transform IR (FTIR) analysis and the X-Ray Diffraction (XRD) analysis have confirmed the presence of sodium bisulfate (NaHSO_4) in the scrubber outlet, but not in the scrubber inlet. This is positive proof that sulfates are generated within the scrubber as the result of oxidation of sodium bisulfite (NaHSO_3) and sodium sulfite (Na_2SO_3) and emitted in the scrubber effluent gas. Also, tests on boilers with flue gas concentrations of 400 to 8,000 ppm SO_2 have shown that there is no correlation between initial

SO₂ concentration and the net sulfate formation rate⁴. This implies that the scrubber has a minimum sulfate emission rate that is virtually unaffected by inlet SO₂ concentration. These data must, however, be evaluated in the context of the potential for significantly increased sulfate loadings to the environment which would result from SO₂ emissions if the boiler flue gases were not controlled.

Based on the analysis of SO₃ and SO₄⁼ emission data, it has been estimated that up to 40 percent of the fine particle emissions at the scrubber outlet could be attributed to scrubber generated NaHSO₄. The remaining portion of the net increase in fine particles across the scrubber may be attributable to the uncertainties associated with the assumptions used in converting the polarizing light microscope number size distribution data to weight size distribution, and to calcium sulfite hemihydrate (CaSO₃·1/2 H₂O) particles generated by the scrubber. This should be confirmed by further study since the unknown or unexplained portion amounts to more than half of the net increase.

Nitrogen Oxides

Mean NO_x emissions during coal firing tests were 421 ng/J (0.98 lb/10⁶ Btu) into the scrubber and 372 ng/J (0.87 lb/10⁶ Btu) out of the scrubber. The comparable data during oil firing were 168 ng/J (0.39 lb/10⁶ Btu) and 161 ng/J (0.37 lb/10⁶ Btu). As expected, little or no NO_x was removed.

Emissions of NO_x varied ±5 percent around these figures. An air leak in the sampling line from the scrubber outlet intermittently allowed sample dilution. Hence, any indicated removal of NO_x by the scrubber is actually a sampling phenomenon, not a real reduction. This was confirmed by later testing.

The NSPS limit is 300 ng/J (0.70 lb/10⁶ Btu) for coal firing and 127 ng/J (0.30 lb/10⁶ Btu) for oil firing, neither of which was met, except at reduced furnace loads.

Carbon Monoxide

Uncontrolled CO emissions averaged 15.9 ng/J (0.04 lb/10⁶ Btu) during coal fired tests and 5.47 ng/J (0.01 lb/10⁶ Btu) during oil firing.

Emissions after scrubbing varied ±5 percent around these figures, with no clear trends depending on test parameters. The analytical sensitivity for CO is about 15 percent of the measured value at these concentrations, hence the small changes across the scrubber are of no significance. Furthermore, an air leak in the sampling line from the scrubber outlet intermittently allowed sample dilution. Thus, any indicated removal of CO by the scrubber is actually a sampling and analytical phenomenon, not a real reduction.

Organics

Uncontrolled hydrocarbon emissions, measured as methane, averaged 5.79 ng/J (0.01 lb/10⁶ Btu) for coal firing and 2.49 ng/J (0.01 lb/10⁶ Btu) for oil firing.

Analytical results from the scrubber outlet are not available due to sample handling problems. Indications from FID, GC, and gravimetric methods for C_1-C_6 , C_7-C_{16} , and $>C_{16}$ organics, respectively, are that, very approximately, 75 percent of organics over C_6 are removed, either in the scrubber or in the gas conditioner in the analytical train.

Polycyclic organic material (POM) was not found in the scrubber inlet or outlet samples from either coal or oil firing at the detection limit of 0.3 mg/m^3 . Since two POM's (benzo-a-pyrene₃ and dibenzo(a,h) anthracene) have been assigned MATE values below the 0.3 mg/m^3 detection limit, additional testing with increased analytical sensitivity is indicated. Also, a more accurate determination of oxygen in the flue gas at the furnace outlet could be important since POM levels decrease as excess air increases at constant temperature.

Inorganics

The emission concentrations during coal firing for 22 major trace elements at the scrubber inlet and outlet are presented in Table 9. To assess the potential degree of hazard of these emissions, the emission concentrations are divided by the Minimum Acute Toxicity Effluent (MATE) values. The MATE values are emission level goals developed under the direction of EPA, and can be considered as concentrations of pollutants in undiluted emission streams that will not adversely affect persons or ecological systems exposed for short periods of time (less than 8 hours)². MATE values for air derived from human health considerations are used as the basis for comparison here.

As shown in Table 9 for coal firing, of the 22 trace elements presented, 17 exceed their MATE values at the scrubber inlet and 5 at the scrubber outlet. The four trace elements in the scrubber outlet that pose a potential hazard are arsenic, chromium, iron, and nickel. Additionally, it may be noted that the emission concentration of beryllium at the scrubber outlet is equal to its MATE value. At this emission concentration, the total beryllium emissions from boilers greater than 50 MW in capacity would amount to more than 10 grams per day and exceed the National Emission Standard for Hazardous Air Pollutants.

In Table 10, the emission factors and the mass emission rates for the 22 major trace elements during coal firing at the scrubber inlet and outlet are presented. The mass emission rates were used to calculate the removal efficiency for these trace elements by the scrubber. The overall removal efficiency for these trace elements is approximately 99.5 percent. As indicated in Table 10, however, some of the trace elements were not removed as effectively as others.

To better understand the removal efficiency of the individual trace elements, the enrichment factor has been computed for each trace element across the scrubber. The enrichment factor is defined here as the ratio of the concentrations of trace element to aluminum in the scrubber outlet, divided by the corresponding ratio in the scrubber inlet. Aluminum is selected as the reference material because it has been shown to partition

TABLE 9 EMISSION CONCENTRATIONS OF TRACE ELEMENTS DURING COAL FIRING

Trace Element	Scrubber Inlet mg/m ³	Scrubber Outlet mg/m ³	MATE Value mg/m ³	Potential Degree of Hazard ^a	
				Scrubber Inlet	Scrubber Outlet
Be ^b	0.1	0.002	0.002	50	1.0
Hg ^c	0.011	0.005	0.05	0.22	0.10
Ca	74	0.036	16	4.6	0.002
Mg	19	0.011	6.0	3.2	0.002
Sb	3.7	0.025	0.050	74	0.5
As	7.8	0.22	0.002	3900	110
B ^b	0.2	0.03	3.1	0.07	0.01
Cd	0.47	0.0010	0.010	47	0.1
Cr	2.6	0.13	0.001	2600	130
Co	3.6	0.012	0.050	72	0.24
Cu	9.6	0.020	0.20	48	0.10
Fe	450	2.4	1.0	450	2.4
Pb	8.5	0.021	0.15	57	0.14
Mn	0.78	0.015	5.0	0.16	0.003
Mo	10	0.027	5.0	2.0	0.005
Ni	1.4	0.063	0.015	93	4.2
V	3.1	0.058	0.50	6.2	0.12
Zn	2.3	0.048	4.0	0.58	0.012
Se	3.2	0.099	0.200	16	0.50
Sr	11	0.058	3.1	3.5	0.019
Al	480	2.6	5.2	92	0.5
Zr	1.6	0.018	5.0	0.32	0.004
Total	1100	6.2			

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

^b Approximate values as determined by Spark Source Mass Spectrometry (SSMS). The other values presented are determined by Inductively Coupled Plasma Optical Emission Spectroscopy (ICPOES).

^c Mercury was determined by cold vapor analysis of SASS train samples.

TABLE 10 EMISSION FACTORS AND MASS EMISSION RATES OF
TRACE ELEMENTS DURING COAL FIRING

Trace Element	Emission Factor, ng/J		Emission Rate, g/hr		Removal Efficiency %	Enrich- ment Factor
	Scrubber Inlet	Scrubber Outlet	Scrubber Inlet	Scrubber Outlet		
Be ^a	0.04	0.001	5	0.09	98	3.7
Hg ^b	0.08	0.037	0.50	0.23	55	84
Ca	32	0.015	3300	1.6	99	0.09
Mg	8.2	0.0046	860	0.48	99	0.11
Sb	1.6	0.010	170	1.1	99	1.2
As	3.4	0.092	350	9.7	97	5.3
B ^a	0.1	0.01	10	1.2	88	2.1
Cd	0.20	0.00042	21	0.044	99	0.4
Cr	1.1	0.054	120	5.7	95	9.5
Co	1.6	0.0050	160	0.53	99	0.6
Cu	4.1	0.0084	430	0.88	99	0.4
Fe	190	1.0	20,000	110	99	0.99
Pb	3.7	0.0088	380	0.92	99	0.5
Mn	0.34	0.0063	35	0.68	98	3.4
Mo	4.3	0.026	450	1.2	99	0.5
Ni	0.60	0.026	61	2.8	95	8.6
V	1.3	0.024	140	2.5	98	3.6
Zn	0.99	0.020	100	2.1	98	3.9
Se	1.4	0.041	140	4.3	97	5.8
Sr	4.7	0.024	500	2.5	99	0.9
Al	210	1.1	22,000	110	99	1.0
Zr	0.69	0.0075	72	0.79	99	2.1
Total	470	2.6	50,000	270	99	

^a Approximate values as determined by SSMS. The other values were determined by ICPOES analysis.

^b Mercury was determined by cold vapor analysis of SASS train samples.

equally among particles of different size*. The enrichment factors presented in Table 10 show that beryllium, mercury, antimony, arsenic, boron, chromium, manganese, nickel, vanadium, zinc, selenium, and zirconium are enriched across the scrubber. The enrichment observed is due primarily to the partitioning of trace elements as a function of particle size, and the greater collection efficiency of the scrubber for the large size particles. It may also be noted that many of the trace elements that show an enrichment trend, such as mercury, selenium and arsenic, either occur as element vapors or form volatile oxides and halides at furnace temperatures. Condensation and surface absorption of the more volatile elements or their oxides and halides downstream of the furnace could, therefore, result in higher concentrations of these trace elements on smaller particles.

Concentrations of 22 major trace elements present in the flue gas during oil firing at the scrubber inlet and outlet are presented in Table 11. MATE values for these elements are also presented for comparison.

Of the 22 elements analyzed, 11 exceed their respective MATE values at the scrubber inlet and 5 exceed their MATE values at the scrubber outlet. The five elements exceeding their MATE values at the scrubber outlet are arsenic, cadmium, chromium, nickel, and vanadium.

Beryllium emissions were measured to be 0.001 mg/m^3 after scrubbing, corresponding to half the MATE value for this element. At this emission concentration, the National Standard for Hazardous Air Pollutants limitation of 10 grams beryllium per day would only be exceeded by boilers of 100 MW capacity or greater.

Emission factors and mass emission rates during oil firing for the 22 elements analyzed are presented in Table 12. Also presented in Table 12 is the scrubber removal efficiency for each element. An overall removal efficiency of 87 percent was obtained for these elements, although several elements were removed with less efficiency, i.e., calcium, arsenic, cadmium, nickel, and vanadium. Note that, with the exception of chromium, all elements that exceeded their MATE values at the scrubber outlet were removed with lower than average efficiency during scrubbing.

The enrichment factor across the scrubber has been computed for each element and is presented in the last column of Table 12.

*Silicon, iron, and scandium have also been used by other investigators as the reference element in the computation of enrichment factors. Notice that iron has no enrichment in this study while silicon and scandium were not measured.

TABLE 11 EMISSION CONCENTRATIONS OF TRACE ELEMENTS DURING OIL FIRING

Element	Scrubber Inlet	Scrubber Outlet	MATE Value	Potential Degree of Hazard ^a	
	mg/m ³	mg/m ³	mg/m ³	Scrubber Inlet	Scrubber Outlet
Be ^b	<0.001	0.001	0.002	<0.50	0.50
Hg ^c	0.0016	0.0002	0.05	0.032	0.004
Ca	0.41	0.070	16	0.026	0.004
Mg	0.31	0.030	6.0	0.052	0.005
Sb	0.062	0.006	0.50	0.124	0.012
As	0.15	0.030	0.002	75.0	15.0
B ^b	0.53	0.039	3.1	0.171	0.013
Cd	0.28	0.066	0.010	28.0	6.60
Cr	0.17	0.018	0.001	170	18.0
Co	0.10	0.012	0.050	2.0	0.24
Cu	0.54	0.007	0.20	2.70	0.035
Fe	4.8	0.28	1.0	4.8	0.28
Pb	0.20	0.013	0.15	1.333	0.087
Mn	0.03	0.004	5.0	0.006	0.001
Mo	0.22	0.025	5.0	0.044	0.005
Ni	1.1	0.20	0.015	73.3	13.33
V	2.7	0.82	0.50	5.40	1.640
Zn	0.61	0.065	4.0	0.153	0.016
Se	0.050	0.006	0.200	0.25	0.03
Sr	0.043	0.001	3.1	0.014	0.0003
Al	5.7	0.48	5.2	1.096	0.092
Zr	0.015	0.001	5.0	0.003	0.0002
Total	18	2.5			

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

^b Beryllium was determined by Spark Source Mass Spectrometry (SSMS). The other values, with the exception of mercury, are determined by Inductively Coupled Plasma Optical Emission Spectroscopy (ICPOES) analysis.

^c Mercury was determined by cold vapor of SASS train samples.

TABLE 12 EMISSION FACTORS AND MASS EMISSION RATES OF
TRACE ELEMENTS DURING OIL FIRING

Element	Emission Factor, ng/J		Emission Rate, g/hr		Removal Efficiency %	Enrichment Factor
	Scrubber Inlet	Scrubber Outlet	Scrubber Inlet	Scrubber Outlet		
Be ^a	<0.0003	0.0003	<0.04	0.04	Unknown	>11.9
Hg ^b	0.0006	0.0001	0.05	0.006	87	1.48
Ca	0.13	0.022	16	2.7	83	2.03
Mg	0.10	0.0094	12	1.1	91	1.15
Sb	0.02	0.0019	2.5	0.23	91	1.15
As	0.049	0.0094	5.9	1.1	81	2.37
B	0.17	0.012	21	1.5	93	0.87
Cd	0.091	0.021	11	2.5	77	2.80
Cr	0.055	0.0057	6.7	0.69	90	1.26
Co	0.033	0.0038	3.9	0.46	89	1.43
Cu	0.18	0.002	21	0.27	99	0.15
Fe	1.6	0.088	190	11	95	0.69
Pb	0.065	0.0041	7.9	0.50	94	0.77
Mn	0.010	0.0013	1.2	0.15	87	1.58
Mo	0.072	0.0079	8.7	0.95	89	1.35
Ni	0.36	0.063	43	7.7	83	2.16
V	0.88	0.26	110	31	71	3.61
Zn	0.20	0.02	24	2.5	90	1.27
Se	0.016	0.002	2.0	0.23	87	1.43
Sr	0.014	0.0003	1.7	0.038	98	0.28
Al	1.9	0.15	220	18	92	1.0
Zr	0.0049	0.0003	0.59	0.038	94	0.79
Total	6.0	0.78	710	96	87	

^a Beryllium was determined by SSMS. The other elements, except mercury, were determined by ICPOES.

^b Mercury was determined by cold vapor analysis of SASS train samples.

Chloride, Fluoride, Nitrate

The results of anion analyses on extracts of particulate matter collected at the inlet and outlet of the scrubber are presented in Table 13. Fluoride is removed with reasonable efficiency, as is to be expected from the overall high removal efficiencies of the trace element cations with which fluoride may be associated. The lower removal efficiency for nitrate suggests that it may be preferentially associated with the fine particles which are not efficiently removed by the scrubber. The chloride removal is much higher in the coal fired tests than in the oil fired, which may be anomalous or may reflect the occurrence of some coal chloride in larger mineral inclusions as contrasted to organic chlorides in oil which would form smaller particles during combustion.

TABLE 13 CHLORIDE, FLUORIDE, AND NITRATE
EMISSIONS FROM COAL AND OIL FIRING

Fuel	Cl ⁻			F ⁻			NO ₃ ⁻		
	Inlet mg/J	Outlet mg/J	Removal Efficiency %	Inlet mg/J	Outlet mg/J	Removal Efficiency %	Inlet mg/J	Outlet mg/J	Removal Efficiency %
Coal	4.7	<0.004	>99	0.22	<0.03	>86	<0.48	<0.25	>52
Oil	0.15	0.072-0.075	52	0.017	0.002-0.003	89	0.076	0.033	57

Solid Emissions

Three solid waste streams are produced by the system:

- o Bottom ash
- o Fly ash
- o Scrubber cake.

Table 14 shows the approximate quantities of bottom ash and scrubber cake that were produced. Only small quantities of fly ash were collected during the test because the multiclone malfunctioned.

The scrubber cake produced after filtration has the characteristics of silty soils, but its behavior closely resembles a clay in many respects. As obtained from the vacuum filter, the scrubber cake consists of small lumps and appears to be relatively dry; however, the water content generally ranges from about 30 to 50 percent.

TABLE 14 GENERATION RATE OF SOLID WASTE FROM
10 MW CONTROLLED INDUSTRIAL BOILER

Waste	Rate of Production, kg/hr	
	Coal Firing	Oil Firing
Bottom ash	80	1
Fly ash	240 ^a	5 ^a
Scrubber cake	700	400

^aThis is the amount of fly ash recovered by the cyclone collector. Approximately 25% of the fly ash is recovered in the scrubber and removed with the scrubber cake.

If it is assumed that calcium sulfite hemihydrate ($\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$) is formed as a result of the SO_2 scrubbing and Na_2SO_3 regeneration processes, then the mass balance of coal firing in Table 15 shows that the scrubber cake is composed of 28.5 percent coal fly ash and 23.8 percent $\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$. However, if the multiclone had not malfunctioned during the test, more fly ash would have been removed upstream of the scrubber and the fly ash content of the scrubber cake would have been lowered proportionately. The amount of scrubber cake produced could be reduced to 600-750 kg/hr on wet basis, assuming approximately 60 to 80 percent multiclone efficiency.

Table 15 also shows the estimated composition of scrubber cake produced during oil firing. The cake is composed of 44 to 50 percent unbound water and at least 47 percent calcium sulfite hemihydrate. These data reflect the low particle emissions which are characteristic of oil firing. Only 1 percent of the scrubber cake during oil firing is estimated to be primary particle emissions, due to their small size and low removal efficiency.

Although the scrubber cake material is composed predominantly of relatively insoluble solids (calcium sulfite, calcium sulfate, and some calcium carbonate), the interstitial water does contain soluble residues of lime, sulfate, sulfite, and chloride salts. Trace elements in the fly ash may also contribute to the leachate from the disposed scrubber cake and are of special concern. The concentrations of 20 trace elements in the scrubber cake are presented in Tables 16 and 17. Note that, except for boron, the trace element concentrations in the scrubber cake from coal firing far exceed the MATE values for solids. Except for antimony, boron, molybdenum, and zinc, all trace elements in oil fired scrubber cake were found to exceed human health based MATE values for solids. Similarly, except for boron, all trace elements were found to exceed ecology based MATE values for solids. These results are a consequence of reducing a high volume of low concentration wastes to a low volume of concentrated wastes. The high potential degree of hazard for most elements appears to warrant disposal of these solid wastes in specially designed disposal areas.

TABLE 15 ESTIMATED SCRUBBER CAKE MASS BALANCE

Component	Contribution to Scrubber Cake			
	kg/hr		Weight %	
	Coal	Oil	Coal	Oil
Fly ash removed by scrubber	324	5	29	1
$\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$ formed from SO_2 scrubbing and Na_2SO_3 regeneration	262	210	24	47
CaSO_4 , CaCO_3 , Na_2SO_3 , $\text{Ca}(\text{OH})_2$, NaHSO_4 , and Na_2SO_4 losses (estimated)	10-85	6-35	1-8	1-8
Water	429-504	193-222	39-46	44-50
Average	1,100	443	100	100

The concentrations of 20 trace elements present in fly ash are shown in Table 18. Again, in almost every case, the trace element concentration in the fly ash far exceeded its MATE value for solids. Trace element concentrations in the bottom ash would be similar to those of the fly ash, except that the more volatile elements and the elements that form volatile compounds would be more enriched in the fly ash. Thus, the concentrations of arsenic, antimony, boron, chromium, manganese, nickel, vanadium, zinc, selenium, and zirconium would all be lower in the bottom ash.

The overall mass balances for the 20 trace elements have been performed and the results are summarized in Tables 19 and 20. The percentage of trace element in the feeds that could be located in the effluent streams (scrubber cake, scrubber effluent gas, bottom ash, and fly ash) is used as a measure of mass balance closure. Except for boron, copper, strontium, and zirconium, the closure of mass balance for the trace elements in coal has been found to be good.

Good mass balance closure for the trace elements in oil was obtained for arsenic, boron, chromium, cobalt, copper, molybdenum, nickel, vanadium, zinc, and selenium. However, as expected due to these extremely low elemental concentrations, mass balance closure for some elements is poor. Instances in which the effluent flow rate of an element substantially exceeded the input feed rates, such as with iron and aluminum, may be the result of the extremely high elemental concentrations attained during coal firing and subsequent contamination of the recycle scrubber solution.

TABLE 16 INORGANIC CONTENT OF SCRUBBER CAKE FROM COAL FIRING (DRY BASIS)

Element	Concentration µg/g	MATE Value, µg/g		Potential Degree of Hazard ^a	
		Health	Ecology	Health	Ecology
Ca	60,715	480	32	126	1,897
Mg	1,458	180	174	8.1	8.4
Sb	315	15	0.4	21	788
As	532	0.5	0.1	1,064	5,320
B	88	93	50	0.9	1.8
Cd	13	0.1	0.002	130	6,500
Cr	141	0.5	0.5	282	282
Co	424	1.5	0.5	283	848
Cu	112	10	0.1	11	1,120
Fe	47,241	3.0	0.5	15,738	94,482
Pb	297	0.5	0.1	594	2,970
Mn	51	0.5	0.2	102	255
Mo	1,117	150	14	7.4	80
Ni	114	0.45	0.02	253	5,700
V	195	5.0	0.3	39	650
Zn	282	50	0.2	5.6	1,410
Se	256	0.10	0.05	2,560	5,120
Sr	642	92	--	7.0	--
Al	45,310	160	2.0	283	22,655
Ar	106	15	--	7.1	--
Total	159,409				

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

TABLE 17 INORGANIC CONTENT OF SCRUBBER CAKE FROM OIL FIRING (DRY BASIS)

Element	Concentration µg/g	MATE Value, µg/g		Potential Degree of Hazard ^a	
		Health	Ecology	Health	Ecology
Ca	200,000	480	32	417	6,250
Mg	3,799	180	174	21	22
Sb	3 ^b	15	0.4	0.2	7.5
As	15 ^b	0.5	0.1	30	150
B	40	93	50	0.4	0.8
Cd	1 ^b	0.1	0.002	10	500
Cr	15	0.5	0.5	30	30
Co	19 ^b	1.5	0.5	13	38
Cu	16	10	0.1	2	160
Fe	2,164	3.0	0.5	721	4,328
Pb	6 ^b	0.5	0.1	12	60
Mn	6	0.5	0.2	32	80
Mo	14 ^b	150	14	0.1	1
Ni	132	0.45	0.02	293	6,600
V	203	5.0	0.3	41	677
Zn	36	50	0.2	0.7	180
Se	9 ^b	0.10	0.05	90	180
Sr	239	92	--	2.6	--
Al	1,684	160	2.0	11	842
Zr	37	15	--	2.5	--
Total	208,450				

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

^b SSMS analyses were utilized where ICPOES analysis provided upper limit data only.

TABLE 18 INORGANIC CONTENT OF FLY ASH FROM COAL FIRING

Element	Concentration $\mu\text{g/g}$	MATE Value, $\mu\text{g/g}$		Potential Degree of Hazard ^a	
		Health	Ecology	Health	Ecology
Ca	378	480	32	0.8	12
Mg	2,478	180	174	14	14
Sb	438	15	0.4	29	1,095
As	1,015	0.5	0.1	2,030	10,150
B	20	93	50	0.2	0.4
Cd	18	0.1	0.002	180	9,000
Cr	434	0.5	0.5	868	868
Co	408	1.5	0.5	272	816
Cu	320	10	0.1	32	3,200
Fe	129,330	3.0	0.5	43,110	258,660
Pb	438	0.5	0.1	876	4,380
Mn	121	0.5	0.2	242	605
Mo	1,288	150	14	9	92
Ni	165	0.45	0.02	367	8,250
V	376	5.0	0.3	75	1,253
Zn	179	5.0	0.2	36	895
Se	378	0.10	0.05	3,780	7,560
Sr	728	92	--	8	--
Al	109,450	160	2.0	684	54,725
Zr	187	15	--	12	--
Total	248,149				

^a Potential degree of hazard is defined as the ratio of the discharge concentration to the MATE value.

TABLE 19 MASS BALANCE ON TRACE ELEMENTS FROM COAL FIRING

Element	Coal Feed g/hr	Scrubber Cake g/hr	Scrubber Effluent Gas g/hr	Bottom and Fly Ash ^a g/hr	Percent Recovery ^b
Ca	2,794	40,072	1.6	30	c
Mg	1,270	962	0.5	198	91
Sb	308	208	1.1	35	79
As	497	351	9.7	81	89
B	8.7	58	1.2	1.6	700
Cd	12.7	8.6	0.04	1.4	79
Cr	174	93	5.7	35	77
Co	461	280	0.53	33	68
Cu	26	74	0.88	26	39
Fe	44,455	31,179	110	10,346	94
Pb	308	196	0.92	35	75
Mn	44	34	0.68	9.7	100
Mo	1,063	737	1.2	103	79
Ni	134	75	2.8	13	68
V	171	151	2.5	30	107
Zn	203	186	2.1	14	100
Se	265	169	4.3	30	77
Sr	247	424	2.5	58	196
Al	50,806	29,905	110	8,756	76
Zr	980	70	0.79	15	9

^a For mass balance calculations, bottom ash has been assumed to have the same trace element concentrations as fly ash. This is an approximate assumption, as some trace elements are enriched in the fly ash.

^b Percent recovery is 100 times the ratio of the sum of the emissions for a trace element to the trace element in the coal feed.

^c Percent recovery is not calculated because most of the calcium in the scrubber cake is from the lime slurry.

TABLE 20 MASS BALANCE OF TRACE ELEMENTS FROM OIL FIRING

Element	Oil Feed g/min	Scrubber Cake g/min	Scrubber Outlet g/min	Percent ^a Recovery
Ca	16.4	50,000	2.8	b
Mg	(12.2) ^c	950	1.2	>1,000
Sb	(2.4)	0.8 ^d	0.2	42
As	(5.9)	3.9 ^d	1.2	68
B	(20.9)	10.0	1.5	55
Cd	(11.0)	0.2 ^d	2.6	25
Cr	3.6	3.9	0.7	125
Co	(3.9)	4.7 ^d	0.5	133
Cu	4.2	4.1	0.3	105
Fe	36.7	541	11.0	>1,000
Pb	(7.9)	1.5 ^d	0.5	25
Mn	(1.2)	4.0	0.2	350
Mo	(8.7)	3.5 ^d	1.0	52
Ni	47.7	33.0	7.9	91
V	108.8	50.7	32.3	76
Zn	8.9	9.1	2.7	133
Se	(2.0)	2.4 ^d	0.2	136
Sr	0.7	59.8	0.04	>1,000
Al	10.4	421	18.9	>1,000
Zr	(0.6)	9.2	0.04	>1,000

^a Percent recovery is 100 times the ratio of its total emission rate (scrubber cake plus scrubber outlet) to its feed rate.

^b Percent recovery is not calculated because most of the calcium in the scrubber cake is from the lime slurry.

^c ICPOES data from the analysis of scrubber inlet particles were utilized when fuel analysis provided upper limit data only.

^d SSMS data were utilized where ICPOES analysis provided upper limit data only.

TABLE 21 ANNUAL EMISSIONS^a

Pollutant	kg/year					
	Scrubber Inlet		Coal/Oil		Scrubber Outlet	
	Coal Firing	Oil Firing	Coal/Oil	Coal Firing	Oil Firing	Coal/Oil
Caseous: NO _x (as NO ₂)	500,810	182,478	2.75	442,520	174,878	2.53
SO ₂	1,127,300	1,006,891	1.12	36,800	27,170	1.36
SO ₃	6,184	8,054	0.77	4,157	5,759	0.72
SO ₄	67,214	23,216	2.90	8,110	9,226	0.88
CO	16,119	5,546	2.91	14,497	5,383	2.69
Organics (as CH ₄)	5,870	2,524	2.32	6,377	2,778	2.30
C ₁ - C ₆ ^b	<5,606	<4,627	--	<5,606	<4,627	--
C ₇ - C ₁₆	345	172	2.00	274	20	13.7
>C ₁₆	2,311	2,646	0.87	335	436	0.77
Total Particulates	2,991,700	59,813	50.0	18,856	15,207	1.24
<1µm	--	--	--	11,691	12,621	0.93
1 - 3µm	--	--	--	5,657	1,824	3.11
3 - 10µm	--	--	--	1,320	704	1.74
>10µm	--	--	--	188	0 ^c	--
Solid: Bottom Ash	~ 778,600	~ 8,444	~ 93	~ 778,600	~ 8,444	~ 93
Fly Ash	~1,800,000	~16,667	~108	~1,800,000	~16,667	~108
Scrubber Cake	0	0	--	8,054,100	3,345,536	2.40

^aAssuming 100% load, 45 weeks per year (7,560 hrs/year).^bRepresents the detection limit of the instrument used.^cRepresents oil firing particulate with a minimum of coal ash contamination.

The scrubber cakes were also analyzed for organics but none was detected. This is expected since concentrations of organics in the flue gas stream were very low.

ANNUAL MULTIMEDIA EMISSIONS

Table 21 presents estimates of the annual emissions of the major pollutants for the controlled and uncontrolled case. It was assumed that the boiler operates at 100 percent load, 87 percent of the year (7560 hours/year).

SCRUBBER EFFICIENCY

Flue gas analyses indicate that the scrubber removes a significant percentage of input sulfur oxides (SO_2 , SO_3 and particulate SO_4), total particulates, and organics of the C_7 class and higher. Scrubber removal efficiency data for these flue gas components are presented in Table 22. As discussed previously, the significance of data indicating NO_x and CO removal appears questionable. Therefore, these components are not included in this discussion. There is no NO_x control equipment.

Average removal efficiencies have been discussed; however, the C_7 and higher hydrocarbons are removed with 77 percent efficiency for coal firing and 85 percent efficiency for oil firing. These fractions comprise 32 to 96 percent of the total generated organics. Hence, based on the total generated organics, a removal efficiency of 25 to 53 percent was obtained for coal firing and 32 to 84 percent for oil firing.

TABLE 22 SCRUBBER EFFICIENCY

	Total Particles	SO_2	SO_3	SO_4^a	Organics	Trace Elements (overall)
Coal	99	97	32	>88	>25	99
Oil	75	97	29	>60	>32	87

^a This removal rate, based on only one data point for each fuel, is actually a net change rate. The scrubber both removes and generates sulfates.

AIR QUALITY

Simplified air quality models were used to estimate the relative ground level air quality resulting from uncontrolled and controlled emissions. Worst case and typical (not average) weather conditions were considered. The worst

case was assumed to be plume trapping, which was assumed to persist for as long as 3 hours. Typical conditions can reasonably be expected to occur almost anywhere in the country. It was further assumed that all species were inert and that no photochemical reactions occurred. Models for particles, SO_2 , NO_x and CO were made. Keeping in mind the assumptions mentioned above, several observations can be made:

- o Controlled emissions of particles for all cases are less than all particle emission standards.

- o For controlled emissions of SO_2 during both coal and oil firing, no standards are exceeded.

- o The NO_x standard is exceeded under both weather conditions during coal firing. During oil firing, the NO_x standard was exceeded under worst case weather conditions but not under typical weather conditions. Since the scrubber does not remove significant amounts of NO_x , there is no substantial difference between the air quality resulting from inlet and outlet emissions. (The boiler has no NO_x controls.)

- o CO standards are not exceeded under any conditions. As with NO_x there is no substantial difference between the inlet and outlet concentrations.

COMPARATIVE ASSESSMENT

Scrubbing removed 99 percent of the particles from coal firing and 75 percent of the particles from oil firing. The lower removal efficiency obtained during oil firing is attributed to the increased fraction of particles smaller than $3\mu\text{m}$; at least 21 percent of the uncontrolled particles from oil firing are less than $3\mu\text{m}$ in diameter while substantially less than 1 percent of uncontrolled particles from coal firing are under $3\mu\text{m}$.

There appears to be a net increase across the scrubber for particles less than $3\mu\text{m}$ in diameter from coal firing. This net increase can be attributed to the poor removal efficiency of the scrubber for fine particles, and to the sodium bisulfate (NaHSO_4) and calcium sulfite hemihydrate ($\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$) particles generated by the scrubber. Both NaHSO_4 and $\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$ have been identified at the scrubber outlet but not at the inlet. Although a very slight increase in particles from oil firing in the $1\text{--}3\mu\text{m}$ range was observed, a net decrease in particles less than $3\mu\text{m}$ was observed during oil firing. Based on the results of coal firing tests, it appears reasonable that scrubber generated particles were present in the scrubber outlet stream during oil firing but that the high fine particle loading associated with oil firing masked detection of these particles. Particle emissions after scrubbing are below either existing or proposed NSPS limits.

Controlled SO_2 emissions for coal and oil firing are lower than either existing or proposed NSPS limitations. The overall uncontrolled sulfur balance indicates that over 92 percent of the fuel sulfur is emitted as SO_2 , less than 1 percent as SO_3 , and approximately 3 percent as SO_4 from coal burning and 1.5 percent from oil firing. The remaining input is in the bottom

ash or is unaccounted for. Sulfates are more efficiently removed than SO_3 (60 percent removal for oil firing and 88 percent for coal firing). This indicates that SO_4 is probably associated with the larger particles, which are more efficiently removed than smaller particles. The higher sulfate removal from the coal flue gases is explained by the higher particulate loading during coal firing.

NO_x emissions increased with increasing load for both coal and oil firing, as expected. Available data indicate that for boiler loadings between 90 and 100 percent, NO_x emissions from coal firing are approximately 3 times greater than from oil firing. Observed reductions of NO_x emissions for coal firing and early oil firing tests appear to be due, at least in part, to air leakage into the scrubber outlet sampling line. Data from later oil firing tests, not known to be subject to leakage problems, indicate that maximum NO_x removal across the scrubber is on the order of 2 percent. Without controls, the emissions of NO_x from clustered coal-fired furnaces will cause the applicable NAAQS to be exceeded.

Uncontrolled CO emissions from coal firing were 3 times those from oil firing, although both emissions are very small. Apparent reductions in CO emissions across the scrubber are not considered significant due to air leakage in the sampling train and the low sensitivity of analysis at the measured CO concentrations.

Organic emissions for coal and oil firing were very similar, and appear to be primarily C_1 to C_6 hydrocarbons and organics heavier than C_{16} . While uncontrolled emission rates for both coal and oil firing are low, emissions of these organics were further reduced by about 85 percent in the scrubber unit. The organic compounds identified in the gas sample from both coal and oil firing were generally not representative of combustion generated organic materials, but were compounds associated with materials used in the sampling equipment and in various analytical procedures. This again confirms the low level of organic emissions. Polycyclic organic material (POM) was not found in the scrubber inlet or outlet at detection limits of $0.3 \mu\text{g}/\text{m}^3$ for either coal or oil firing. MATE values for most POM's are greater than this detection limit. However, since the MATE values for at least two POM compounds -- benzo(a)pyrene and dibenz(a,h)anthracene -- are less than $0.3 \mu\text{g}/\text{m}^3$, additional GC/MS analyses at higher sensitivity would be required to conclusively determine the presence of all POM's at MATE levels. Also a more accurate determination of oxygen in the flue gas at the furnace outlet could be important since POM levels decrease as excess air increases at constant temperature.

The air concentration of trace elements from plant clusters is expected to be approximately 4 orders of magnitude below the "allowed exposure levels" proposed for hazardous waste management facilities.⁵ They are also below typical urban ambient background, except for cobalt and selenium, which approach or slightly exceed endogenous levels. The concentrations are similar from coal and oil firing, except for cadmium, which is 40 times larger from oil firing than from coal firing.

Trace element concentrations in runoff water, which arise from deposition of emissions on soil and foliage, may be about 10^5 times the standards for livestock drinking and potable water. Concentrations due to oil firing are slightly lower than those due to coal firing; however, selenium and molybdenum concentrations in water are predicted to exceed their background levels. Mass closure for most trace elements from coal-firing has been found to be in the 75 to 107 percent range. Mass closure for half of the trace elements from oil-firing is in the 50 to 136 percent range; closure for the remainder of oil-firing trace elements is poorer due to the extremely low elemental concentrations measured and/or contamination of the recycle scrubber solution during coal firing tests. These good closures instill confidence in the validity of the sampling and analysis data. Beryllium emissions after scrubbing were less than or equal to the beryllium MATE value during coal and oil firing. At the measured emission concentrations, the National Standard for Hazardous Air Pollutants limitation of 10 grams beryllium per day would only be exceeded by boilers of 50 MW capacity for coal firing and 100 MW capacity for oil firing.

Chlorides were removed with greater than 99 percent efficiency from coal flue gases and with about 51 percent efficiency from oil flue gases. This difference was attributed to the higher removal efficiency for the larger coal particles. Fluorides were removed with greater than 86 percent and about 87 percent efficiency for coal and oil firing, respectively. Nitrate emissions were removed from coal flue gases with at least 52 percent efficiency and from oil flue gases with 57 percent efficiency.

Scrubber cake production during coal firing was 3.3 times greater than during oil firing. If the multiclone had not malfunctioned, this ratio would have been reduced to 2.7, assuming 60 percent multiclone efficiency. Available data indicate that the principal difference between scrubber cake production rates from coal and oil firing is the particle loading and associated unbound moisture.

The scrubber cake produced from coal firing contains a significant amount of fly ash. Except for boron, trace element concentrations in the scrubber cake have exceeded their MATE values. This is the result of transferring an air pollution problem by scrubbing to an easier solid waste disposal problem. Because the trace elements may contribute to the leachate from the disposed scrubber cake, these solid wastes must be disposed of in specially designed landfills. In such landfills, leachate impact on ground water is expected to be insignificant.

CONCLUSIONS

General

Several major conclusions have evolved from the environmental analysis.

The difference in environmental insult expected to result between coal and oil combustion emissions from a single controlled 10 MW industrial boiler is insignificant. This is because: 1) there are only slight differences in the emissions levels of the pollutants, 2) the absolute impact of either fuel use is insignificant, and (3) the effectiveness of the control equipment makes environmental impacts small.

The environmental impacts of emissions from a cluster of five 10 MW industrial boilers at 200-meter intervals aligned with the prevailing wind are potentially significant. The impacts include health effects, material damage, and ecological effects from high ambient levels of SO₂, NO_x, and suspended particulate matter; health effects and ecological damage due to trace metal accumulation in soils and plants; and degradation due to visibility reduction and the presence of waste disposal sites.

The environmental acceptability of a cluster of controlled industrial boiler emissions depends more on site specific factors (e.g., background pollution levels, location and number of other sources) than on the type of fuel used. Careful control of the site-specific factors can avert potential environmental damage and generally compensate for any differential effects arising between the use of coal or oil.

With the possible exception of ambient levels of NO_x, the risk of violating the NAAQS due to operation of clusters of controlled industrial boilers is essentially the same whether the fuel is coal or oil. Based on tests of the reference 10 MW boiler (which was not controlled for NO_x emissions), localized NO_x concentrations produced by coal firing are estimated to be twice those resulting from oil firing, and greater than those permitted by the NAAQS for 24 hour and 1 year averaging periods.

Coal firing appears to produce a greater enrichment of trace elements in the flue gas desulfurization filter cake than does oil firing; however, the scrubber cake resulting from either coal or oil firing contains sufficient amounts of heavy metals and toxic substances to require specially designed disposal areas.

Health Effects

Regional emission levels of suspended sulfates from controlled oil or coal fired industrial boilers would not be expected to cause a significant impact on regional health.

Sulfate emissions from clusters of controlled industrial boilers might be expected to cause significant adverse health effects in a localized area near the plant cluster. Oil firing would be expected to result in localized health effects about one-third less severe than those resulting from coal firing since oil firing produces only one-third the particle emission of coal firing.

The impact of solid waste generation on health is essentially the same for controlled coal firing and oil firing, if suitable land disposal techniques are employed to ensure sufficiently low leaching rates and migration of trace elements to groundwater and the terrestrial environment.

The concentration of metals in runoff waters due to controlled oil firing is predicted to be slightly less than that occurring from controlled coal firing; in either case, hazard to human health by drinking water is remote.

Trace element emissions from clusters of controlled industrial boilers may significantly increase local background levels in drinking water, plant tissue, soil, and the atmosphere; however, the expected increases in the levels of such elements are generally several orders of magnitude less than allowable exposure levels. Oil firing is estimated to cause cadmium burdens in plants approaching levels injurious to man. Because cigarettes contain significant cadmium levels, smokers are more apt to achieve thresholds of observable symptoms for cadmium exposure if they also consume additional cadmium via the food chain. Coal firing may produce plant concentrations of molybdenum which are injurious to cattle.

Ecological Effects

The potential for crop damage from either controlled coal firing or oil firing depends greatly on ambient levels of NO_x , SO_2 , or trace element soil concentrations. If such levels are currently high, localized plant damage would be expected to occur within 1 to 2 km of a controlled boiler cluster. Leaf destruction from SO_2 exposure would be expected to be slightly more severe in the vicinity of a cluster of controlled boilers which are coal fired as opposed to oil fired. Plant damage may possibly occur even at levels below ambient air standards.* For uncontrolled NO_x emissions, plant damage would be expected to be significantly greater in the vicinity of the coal fired cluster, because of the higher levels of ambient NO_x produced. Emissions of CO and hydrocarbons will have negligible impacts on plants. The likelihood of damage occurring in plants due to emissions of trace elements from either controlled oil or coal firing is remote, with the possible exception of injury due to elevated levels of molybdenum and cadmium in plant tissue resulting from coal firing and oil firing, respectively.

The impact of fossil fuel combustion in controlled oil or coal fired boilers on plant damage via acid precipitation would be relatively insignificant. The levels of suspended sulfate (the precursor of acid rain) would be essentially the same whether the controlled boilers are coal or oil fired.

Measurements and analyses of leaching rates at experimental solid waste disposal sites indicate that landfills of untreated flue gas desulfurization system scrubber cake can be constructed without significant adverse impacts.

Societal Effects

The impact of boiler emissions on corrosion in the local area near a cluster of controlled industrial boilers will be significant⁶. The corrosion rate will be slightly greater when the boilers are coal fired; however, the extent of this overall impact (oil or coal) is minor compared to that which occurs when industrial boilers are uncontrolled.

* See pp. 5-29 to 5-32 of volume 2 of reference 3.

The increase in annual total suspended particulate matter and the resulting soiling damage in the vicinity of a cluster of controlled industrial boilers results in additional cleaning and maintenance costs about 10₇ to 15 percent greater than that already experienced in a typical urban area⁷. The cleaning costs may be slightly greater when the boilers are coal fired.

Emissions of particulate matter from controlled industrial boilers will result in visibility reduction. This form of environmental degradation will occur in a localized area near the boiler cluster, and occurs to essentially the same extent whether the controlled boilers are oil or coal fired.

Total land disposal requirements for scrubber waste generated by controlled coal firing are 3 times greater than those for controlled oil firing. Disposal of the scrubber wastes may result in significant depreciation of property value and environmental degradation in the area of the disposal site. These impacts would be more severe if boilers use coal rather than oil.

Economic Effects

The direct economic impacts associated with residuals of fuel combustion involve the costs of damages (or benefits) sustained when the residuals enter the environment. Second order economic impacts associated with the residuals involve the alterations that occur in employment, the tax base, energy prices, income, and land values due to the damages (or benefits) resulting from combustion residuals. The quantification of direct economic impacts involves the difficult task of ascribing economic values to environmental changes. Quantification of second order economic effects is yet more difficult because of gaps in knowledge which make it impossible to determine the complex relationships between cost and the numerous socioeconomic factors involved. The scope and data of this program did not permit such quantification. Because the significant effects of direct economic impacts occurring from controlled oil fired and coal fired boiler emissions are limited to a relatively small area near the source, the total costs of the incremental environmental damages are apt to be insignificant on the regional basis. Consequently, significant incremental second order economic impacts (such as changes in hospital employment, alteration of tax bases, or changes in income) would be unnoticeable between controlled oil and coal fired industrial boilers.

Energy Effects

Obviously, our abundance of coal and the uncertainty of our oil supplies are the driving forces for studies such as the coal versus oil comparative assessment study. Whether coal assumes a more significant role as an energy source by national choice or because we no longer have a choice, it is essential to be aware of and to be prepared to deal effectively with any environmental problems which result. Thus, the comparative impact of coal versus oil firing is indeed complex and involves consideration of all aspects of energy supply and use, including emissions characterization, multimedia environmental impacts identification, comparison of projected impacts with accepted levels of impacts, and evaluation of techniques for mitigating

unacceptable levels of impacts. This study, which is one of several projects in the CCEA Program, is intended as a prototype of future projects dealing with the impacts and control of stationary conventional combustion processes.

While this study identified some potentially significant differences between coal and oil firing in clusters of boilers, the fuel choice of oil or coal may be a relatively minor factor in determining the environmental acceptability of controlled industrial boilers. Other site specific and plant design factors may exert greater environmental effects than fuel choice. As concern for environmental protection increases, the issue may become whether the increasing use of fossil fuels can be continued at the present levels of control technology without potential long term dangers. If it is found that long term effects of pollution (e.g., trace metals accumulation, lake acidity from acid rains) from fossil fuels combustion and other sources are environmentally unacceptable, it is clear that energy use may be affected. Increasing control requirements could result in energy cost increases to the level where the combustion of fossil fuels loses its economic advantage over other, cleaner sources of energy production.

ACKNOWLEDGEMENT

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OPERATING HISTORY AND PRESENT STATUS
of the
GENERAL MOTORS DOUBLE ALKALI SO₂ CONTROL SYSTEM

Thomas O. Mason, Chevrolet Div., GMC, Parma, Ohio
(Speaker)

Edward R. Bangel, Chevrolet Div., C.O., GMC, Warren, Michigan

Edmund J. Piasecki, EAS, GMC, Warren, Michigan

Robert J. Phillips, EAS, GMC, Warren, Michigan

I. INTRODUCTION

The General Motors version of the double-alkali wet scrubbing process went on line in March of 1974 at the Chevrolet Motor Division facility in Parma, Ohio. This installation, was at that time, the latest development in a program started by General Motors engineers in 1968 to find an environmentally acceptable means of burning coal in an industrial-size power plant. Most of the boilers in the Corporation range from steaming capacities of 50,000 to 150,000 pounds per hour, with an average plant steaming capacity of 250,000 pounds per hour. Generally, the pressure ranges below 200 psi, and the steam is used in process operations and building heat.

There are a number of operating characteristics associated with industrial boilers which have to be considered in the scrubber system design. These include boiler type, load fluctuation, high excess air, and dust load. In addition to these basic design characteristics, the system had to be simple to operate, reliable, and economically competitive with other control alternatives.

With this in mind, the Chevrolet Parma, Ohio, plant was selected in 1972 for a prototype wet scrubbing system. The history and status of the installation follow.

II. DESCRIPTIVE LAYOUT OF THE FACILITY AND EQUIPMENT

The Chevrolet Parma facility includes a sheet metal stamping plant, an automatic transmission manufacturing plant, and a propeller drive shaft manufacturing plant. Steam is supplied for process parts washers and building heat from four boilers, centrally located in a separated facility. Two, 60,000 pound-per-hour boilers were installed in 1948 and two, 100,000 pound-per-hour boilers were installed in 1966 and 1967. The boilers are all traveling grate, continuous-ash-dump type with stoker feeders. They are also equipped with mechanical dust collectors. The larger boilers have economizers, and stack discharge temperatures are approximately 350 degrees Fahrenheit. The stack discharge temperature of the smaller boilers is approximately 600 degrees Fahrenheit.

The double-alkali type system was chosen because of its potential reliability over lime/limestone systems. The system is closed-loop with scrubbing being accomplished by use of a dilute caustic mixture and regeneration with lime. Each boiler has its own scrubber and caustic control. The lime regeneration system is common to all units.

Additional draft fans were necessary to overcome the seven to eight inch pressure drop of the scrubbers. These fans are

installed upstream of the scrubbers to avoid corrosion problems.

The scrubbers have three bubble cap absorption trays and a mesh-type mist eliminator. This type of scrubber was chosen to obtain the highest sulfur dioxide removal along with maximum turndown potential. All wetted surfaces are constructed of 316L stainless steel.

Caustic make-up to the scrubbers is determined by the pH of the liquor as it is recirculated. Liquid blowdown is diverted to two in-line reactor tanks where slaked lime slurry is added. These tanks overflow to a reactor clarifier where solids settling occurs and the lime reaction continues. The overflow from this clarifier goes to a second clarifier, where soda ash is added for softening and sodium make-up, and additional solids settling occurs. Overflow from the second clarifier is the make-up liquor used to replace blowdown at the scrubber.

III. ASSIGNMENT OF MANPOWER

It was determined that scrubber operation was to be placed under the supervision of the powerhouse personnel. This was done in order to coordinate the scrubber operations with steam generation.

A rotating continuous operation was established with boiler operators operating the scrubber. Union agreements limit us to the use of powerhouse personnel to run the scrubber facility. Although these people normally run water tests, set valves, monitor controls, and adjust set points in their boiler duties, they have a limited chemical processing background. This has created some operational difficulties.

There are a total of eight hourly people permanently assigned to the operation. This is compared to sixteen hourly people already employed for powerhouse operations. Of these eight, four are assigned as shift operators, two to vacuum filtration and sludge handling, and two to maintenance activities. No additional supervision is assigned at the scrubber facility at this time.

A set of operational guidelines and instruction books were published. Personnel were indoctrinated in classes and

assigned duties and shifts. With no other system to study, we decided to go into operation without further delay and work through our problems.

IV. EARLY DEVELOPMENTS

The scrubbers were started up for the first time on February 28, 1974. A number of initial problems were encountered, many of these were due to lack of experience with the system. For example, just prior to an open house, we experienced an upset when a maintenance employe started cleaning the building walls with a soap solution. Since all spills or leaks drain back into the system via sump pumps, the soap entered the process and caused all the solids in the clarifiers to go into suspension. During our open house, we were scrubbing with slurry. This eventually shut down the system.

We also encountered a number of functional problems with the system. First, we had a problem with reliable pH control. This was due to initial location of the pH sensor. Since the probe was in a pressure line, it would frequently break. In addition, because the response time was so short, pH readings were unreliable. We tried to solve the problem by placing the probe in a gravity sample line off the top tray. This stopped the breakage, but we soon coated the entire probe with solids.

These experiences led to the present arrangement of sampling pH in a gravity sample line off the bottom tray. This lower pH location has eliminated the scaling problem and allowed us

to maintain reliable pH control, which is maintained at a level of pH six.

Eventually we experienced a problem due to the lack of intermixing of recycle liquor and scrubber feed which caused the top tray to plug. This was the result of a high pH condition on the top tray, with calcium carbonate scaling shutting down operations within eight hours. An early fix was to have both lines feed into a mixing box above the top tray of the scrubber. This proved ineffective because sufficient time was not allowed for mixing. The next step was to put caustic feed into the recycle tank. This effectively reduced top tray pH and eliminated the carbonate scaling problem. Unfortunately, solids build-up of calcium sulfite occurred at a fantastic rate in the recycle tanks, causing them to fill with solids. This was primarily due to the long hold-up time in the tank at a lower pH. Finally, we determined that in order to minimize scaling, the proper place to introduce caustic make-up was in the recycle line just prior to entering into the top tray of the scrubber. This has greatly reduced the scrubber scaling problem.

An additional problem was encountered with settling calcium salts in the first clarifier. The problem was caused by hindered settling due to excess sludge recycling and trouble maintaining a constant sludge blanket in the bottom of the

clarifier. This was resolved by eliminating sludge recycle and by not operating with a sludge blanket.

We have had a very long learning curve on the operation of vacuum filters. The first cloth was a polypropylene material that had good wear characteristics but poor cake release. This was replaced with nylon and improvements have resulted.

Also contributing to the problem of cake release was the high percentage of calcium sulfite in the cake. Cake characteristics were greatly improved by adding an oxidation step prior to lime addition. This produces a higher percentage of calcium sulfate. We are now obtaining approximately 60 percent solids in the cake.

Overdesign of the system has caused problems with control. The system was originally designed to handle 400,000 lbs./hr. of steaming capacity burning three percent sulfur coal. However, the maximum actual operations have been only fifty percent of design capacity. Since the controls were designed for maximum levels, there have been problems measuring flows and controlling the operation at these reduced conditions. This made it necessary to reduce line sizes in specific areas of the process in order to more accurately measure and control liquid flows.

We also found a serious problem with closed gravity lines carrying solids to the clarifiers. This caused numerous outages of the system due to plugging of these lines. We finally resolved the problem by installing removable covered troughs of a rectangular cross section which increased the flow rate in these lines and made them accessible for maintenance.

V. CURRENT OPERATING PROBLEMS

We are currently in the process of modifying our lime feed systems. The original system has never been able to accurately control the amount of lime added to the system. This was due to the number of variables in the control loop which made this lime addition so complex it could not be maintained. We are simplifying the system by installing linear feed pumps and proportioning that feed to caustic flow to the scrubbers.

Mist carryover continues to be one of our most serious problems. Initial scrubber design did not allow for utilization of the entire mesh pad. This was corrected by modifying the transition from the scrubber to the exhaust stack and by relocating the mesh mist eliminator. However, this has not totally resolved the problem. We are continuing to investigate potential solutions.

We are also currently modifying the vacuum filter agitators, slurry inlet, and cake support roll. We hope to have better solids distribution over cloth and improve cake filtering characteristics.

VI. SYSTEM PERFORMANCE VERSUS TIME AND OPERABILITY

The scrubbers have held up very well with little or no corrosion discernible. The rubber-lined recycle pumps have proved to be very maintenance free. The recycle tanks are showing corrosion and will require relining. They are mild steel painted with a bitumastic coating. We are presently patching them and relining with plastic. Corrosion of the stacks occurred almost immediately after startup. A stainless liner was added within the first year, and that has also failed. We have since replaced them with FRP stacks.

The sludge handling operation receives the most abuse and, therefore, requires the most service. The present sludge pumps, however, have provided good service. The original sump pumps failed early and were replaced with air operated diaphragm pumps.

Piping in the system has been mostly trouble free. All piping in areas of the system where corrosion could occur is hand-laid, fiber-cast plastic.

The ratio of scrubber hours versus boiler hours has been increasing to a level now near 70 percent. One of the factors that must not be overlooked in this evaluation is that these

hours include many scheduled shut-down periods for planned maintenance, evaluation programs, or system modifications. Typically, the system operates for extended periods of time without interruption.

VII. RELATIVE COST

The latest costs of constructing and operating our system are as follows:

	<u>1978 \$/ton of coal</u>
Capital Charge	6.85
Chemicals	1.26
Utilities	0.93
Solid Waste	0.70
Maintenance	6.21
Labor	<u>4.24</u>
Total	\$20.19

The construction cost to General Motors was 3.2 million dollars, based upon 1973 dollars. This figure includes all design and construction-related costs performed by other corporate divisions.

VIII. FUTURE OUTLOOK

The system is running now. It may require more modifications to reach a reasonable reliability level. We are experiencing carryover problems, and the mist eliminator will have to be modified. This is a major concern, especially in the appearance of the surrounding area.

General Motors continues to believe that until the system's reliability has been improved to 90 percent, this double-alkali process should be considered developmental. With the implementation of the programs outlined, we hope to obtain this goal.

Ladies and gentlemen, it's very difficult to squeeze five years into a short talk, but I thank you for the opportunity to try.

R-C/BAHCO FOR COMBINED SO₂ AND PARTICULATE CONTROL

NICHOLAS J. STEVENS

RESEARCH-COTTRELL, INC., SOMERVILLE, NEW JERSEY

ABSTRACT

The R-C/Bahco system controls SO₂ and particulate emissions in industrial applications where high performance is required. Research-Cottrell started up the system successfully on its first commercial application at Rickenbacker Air Force Base, Columbus, Ohio in March, 1976. The scrubbing system has been operating at high availability (>90%) since the Air Force took beneficial occupancy in September, 1976. An EPA-sponsored test program was conducted to evaluate the viability of the R-C/Bahco technology for air pollution control. Effective SO₂ removal is accomplished using either a lime or limestone reagent. The Air Force switched from lime to limestone in May, 1977 because of the substantial reagent cost savings. High particulate collection efficiencies (98+%) are obtained for fly ash particle sizes above 1 to 2 microns. This paper describes the R-C/Bahco system installed at RAFB, evaluates its performance and operation and indicates costs and range of R-C/Bahco application for coal-fired industrial boilers.

R-C/BAHCO FOR COMBINED SO₂ AND PARTICULATE CONTROL

INTRODUCTION

In September 1974, the United States Air Force took a significant step to demonstrate the applicability of flue gas scrubbing technology to industrial coal-fired plants. Research-Cottrell was awarded a contract for an SO₂ and particulate emissions control system for the Central Heat Plant at Rickenbacker Air Force Base (RAFB) near Columbus, Ohio. An R-C/Bahco scrubber system was selected for this project. The scrubber, which simultaneously accomplishes SO₂ and particulate removal, is based on technology developed by A.B. Bahco in Sweden and tested worldwide on oil fired boilers, incinerators and other applications. The system installed at RAFB, represents the first application of the Bahco system in the U.S., and the first application worldwide on a coal-fired industrial boiler.

The U.S. Air Force awarded a contract to Research-Cottrell in April, 1975, by means of an interagency grant, to evaluate the viability of the R-C/Bahco air pollution control technology. The objective of the EPA sponsored program was to characterize the scrubbing system at RAFB in terms of its performance, reliability and economics for SO₂ and particulate control on industrial-scale, coal-fired boilers.

The R-C/Bahco system is especially effective in those industrial applications where high performance for SO₂ and particulate removal is required. However, the process also has been applied to a wide range of industrial pollution control problems at 18 installations in Japan and Sweden (1). This paper describes the R-C/Bahco system installed at RAFB, evaluates its performance and operation and indicates the costs and range of R-C/Bahco application for coal-fired industrial boilers.

RICKENBACKER AFB HEAT PLANT

The Central Heat Plant at Rickenbacker Air Force Base consists of a bank of six coal-fired hot water generators. The total installed capacity of the stoker-fired boilers is about 330 million BTU/hr. firing rate and they typically burn 11,300 BTU/lb. coal containing 2.5-3.5% sulfur.

The generators emit approximately 6.2 lbs. of SO₂ and 5.6 lbs. of particulate per million BTU of coal burned. The Ohio Air Pollution Control Regulations in force limited emissions to 2.2 lbs. sulfur dioxide and 0.16 lbs. particulate per million BTU of coal burned. Thus, in order to comply with the Ohio Regulations the SO₂ and particulate emissions had to be reduced by 65% and 97.5%, respectively.

The R-C/Bahco system (Figure 1) was designed to treat up to 108,000 ACFM of flue gas generated at the peak winter firing rate of approximately 200 million BTU/hr. The system, which has essentially unlimited turndown capabilities, handles seasonal load variations from 20 to 200 million BTU/hr., SO₂ concentrations up to 2000 ppm and particulate loadings of up to 2 gr/SCFD. In addition, the scrubbing system copes with 100% load increases occurring in as short a time as one hour.

PROCESS DESCRIPTION (Figure 2)

Hot flue gas from each of the Heat Plant Generators is passed into a common flue which contains a by-pass stack. The by-pass stack allows makeup air to be drawn into the system at low load to maintain efficient operation of the mechanical collector and scrubber.

The booster fan forces the flue gas into the first stage of the scrubber where it is vigorously mixed with the scrubbing slurry in an inverted venturi. In this stage, the flue gas is cooled to its adiabatic saturation temperature and SO₂ and particulate are scrubbed from the gas. The partially scrubbed gas rises to the second stage and is contacted with slurry containing fresh reagent to complete the required SO₂ and particulate removal. Gas from the second scrubber stage slurry is separated by centrifugal force to produce an essentially droplet-free effluent.

At Rickenbacker pebble lime is slaked and added directly to the spent slurry recycled to the dissolver tank. The resulting fresh lime mixture is pumped to the second stage (top) venturi to treat the flue gas stream. The slurry flows by gravity from the top stage to the lower stage where it contacts the entering hot flue gas. This countercurrent flow arrangement in the scrubber produces high SO₂ removal combined with efficient reagent usage.

Spent slurry flows by gravity from the lower stage of the scrubber to the dissolver tank. Part of the spent stream leaving the lower stage is diverted to the thickener where the slurry is concentrated to approximately 40% solids. The

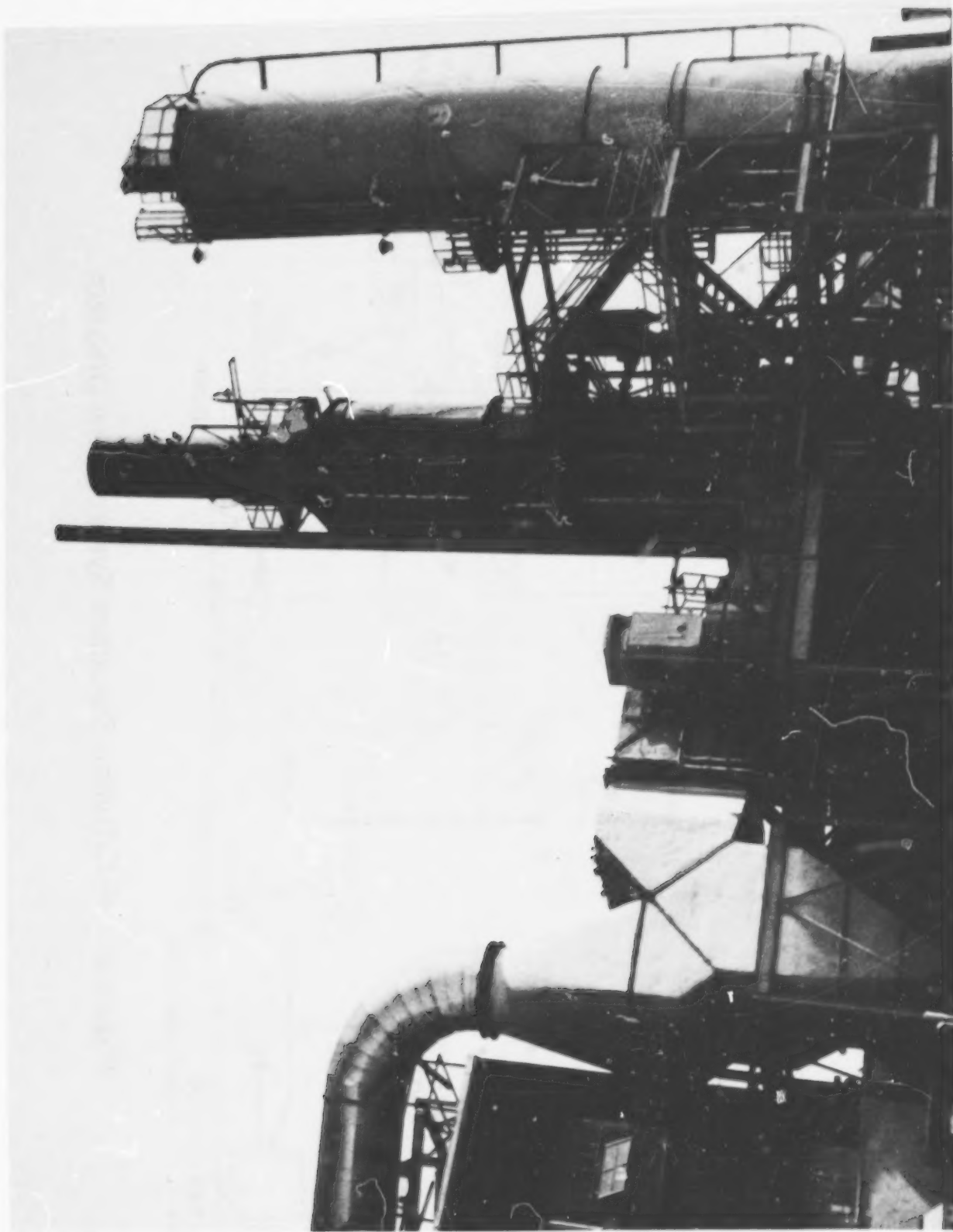


FIGURE 1: The R-C/Bahco Scrubbing System at RAFB

thickener overflow is recycled to the dissolver tank and the thickened sludge is pumped to a Hypalon-lined pond adjacent to the Heat Plant.

MAJOR EQUIPMENT

The major equipment installed in the R-C/Bahco system at RAFB, shown in Figure 2, is described below.

Flue System

The flue system includes individual tie-ins to each of eight boilers. Manual diversion dampers allow gas flow into the flue system or bypassing through individual stacks. In addition, a stack in the main flue upstream from the mechanical collector permits the addition of makeup air to the gas or by-passing of the scrubber.

Mechanical Collector

A Flex-Kleen mechanical collector handles 108,000 ACFM of flue gas at 475°F at dust loads up to 2 gr. per SCF. This collector is located in the main flue upstream from the booster fan. The collector operates at approximately 5 in. W.C. pressure drop at full load and 1.5 in. W.C. at minimum load.

Booster Fan

The booster fan draws gas and air mixtures through the mechanical collector and forces them into the R-C/Bahco scrubber. The fan was oversized by 200 H.P., for a total of 700 H.P., to allow for high gas flow rates at pressures up to 30 in. W.C. for the EPA test program (2). The scrubber normally operates at 15 to 18 in. W.C.

Reagent System

The reagent system consists of a reagent unloading system, storage bin, feeder, and lime slaker. The equipment can handle pebble lime as well as limestone as reagent for SO₂ removal.

The pneumatic unloading system handles 3/4" pebble lime at a maximum rate of 1 ton per hour. The 120-ton capacity storage bin is equipped with a motor driven live bottom which is activated by the lime slaker.

The weigh belt feeder is fitted with manual and automatic controls and a totalizer. Either the SO₂ mass flow rate or dissolving tank pH can be used to control the reagent feed rate. The lime slaker includes a water totalizer, grit removal circuit, high temperature alarm, and dust and vapor venting system.

Lime Dissolving Tank

The lime dissolving tank, which serves as the surge tank for the entire system, is made of 316L SS. In this tank, slaked lime or limestone is blended with spent slurry for recirculation to the scrubber.

R-C/Bahco Scrubber

The scrubber is an R-C/Bahco size 50 module fabricated from 316L SS. This module is approximately twelve feet in diameter by sixty feet high and has a nominal gas handling capacity of 50,000 SCFM. A fiberglass reinforced polyester (FRP) stack, 5.5 feet in diameter and 20 feet high, is mounted on top of the scrubber.

The scrubber module incorporates two inverted fixed-diameter venturis. Each has a level tank located outside the shell of the scrubber with a manually adjustable weir. A fluid mill, which grinds coarse reagent particles, is located in the bottom of the scrubber module.

Second Stage Slurry Recycle Pump

The second stage recycle pump circulates slurry through the entire scrubbing system. The pump is rubber-lined and is rated at 2600 gpm at 20 psig. A 316L SS shaft sleeve and a water purge in the stuffing box are used to minimize wear and corrosion.

Mill Pump

The mill pump is identical to the second stage slurry pump but it operates at 2000 gpm at 25 psig. This pump is the prime mover for the fluid mill at the base of the scrubber module.

Thickener

The thickener, which is 25 feet in diameter and 8 feet in height, is used for solids surge capacity, slurry density control, and thickening sludge for disposal. The tank is Douglas Fir and the rake mechanism is rubber-covered carbon steel. The rake mechanism has a lifting device and torque sensor with a high torque alarm and cutoff for its protection.

Sludge Transfer System

The sludge transfer system includes two thickener underflow pumps and two transfer lines to provide 100% standby capacity. The two underflow pumps are air operated and fitted with replaceable neoprene diaphragms and 316 SS wetted parts. The pumps are capable of pumping up to 40 gpm of sludge at 75 psig. The lines to the sludge pond run underground inside a sleeve to permit easy removal for cleaning or replacement if it is necessary.

Sludge Pond

The sludge disposal pond, shown in Figure 3, is located approximately 700 feet from the scrubbing system. The pond is lined with Hypalon (chloro-sulfonated polyethylene), and is approximately 450 feet long, 250 feet wide and 12 feet deep. An underdrain system allows ground water to be removed from beneath the liner and also serves as a means of detecting any leaks which may occur. The pond was designed to hold sludge produced by scrubbing flue gas from the combustion of 200,000 tons of 5% sulfur coal.

PROCESS CONTROL

A combination of manual and automatic controls are used to adjust and regulate the scrubber system variables.

Manual Controls

The manual mode is used to control variables which do not have to be adjusted frequently or, once set, are essentially constant over the entire operating range of the system. Gas flow rate, slurry circulation rate and first and second stage pressure drops are controlled manually.

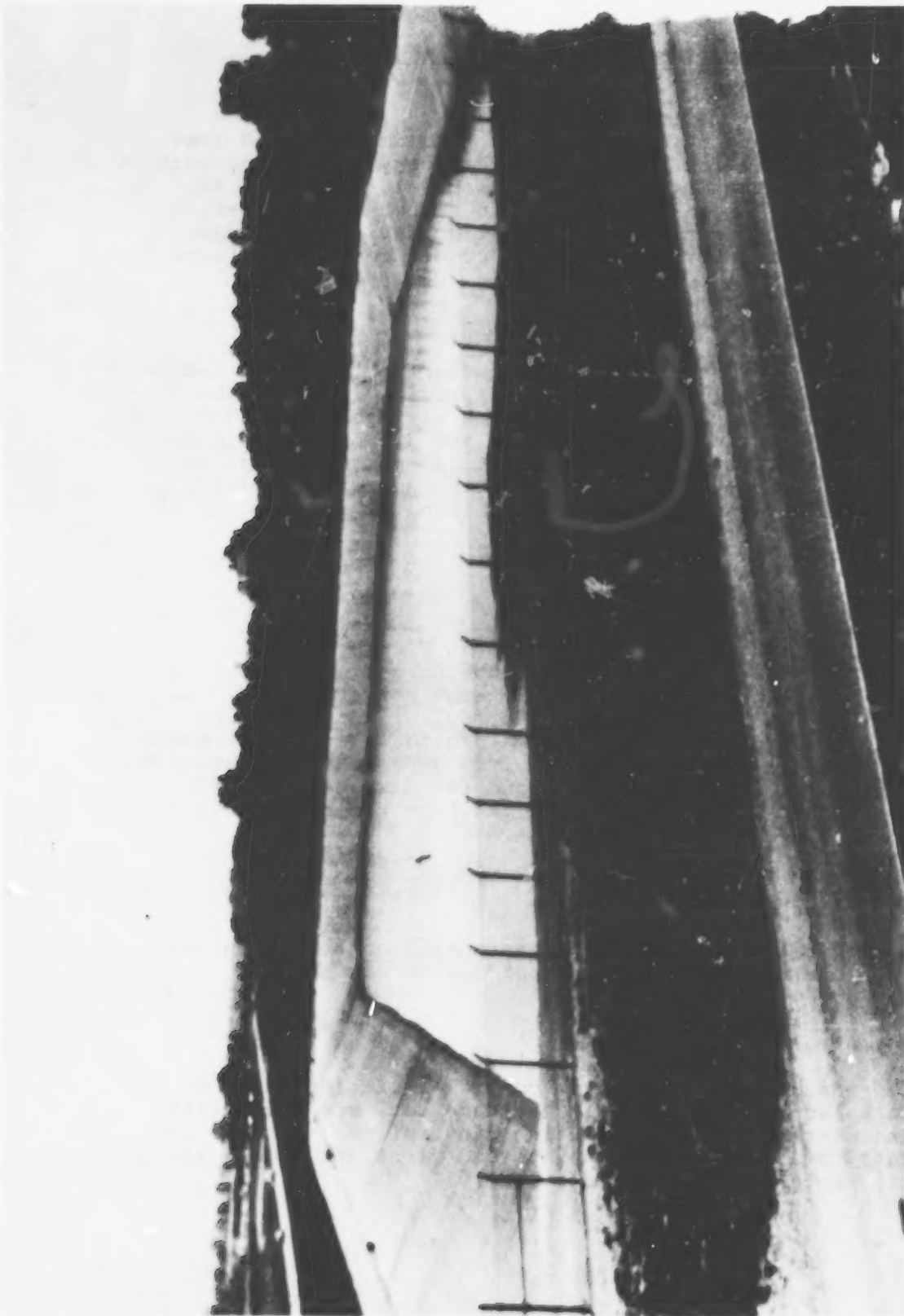


FIGURE 3: Lined Storage Pond at RAFB

Gas Flow Rate At RAFB the gas rate to the scrubber is set manually by adjusting the booster fan inlet dampers. Merely positioning the damper to obtain the desired total gas flow rate is sufficient for control purposes. Variations in flue gas rates are accommodated by mixing makeup air with the flue gas to maintain the desired total gas flow rate. Control is maintained as long as the flue gas volume is less than the total flow that the booster fan is able to accommodate for the damper position selected. If the flue gas flow exceeds this total rate, gas bypasses through the makeup air stack and activates a temperature alarm to alert the Heat Plant operators.

Slurry Circulation Rates All slurry circulation rates are manually adjusted and are set to maintain line velocities between 4 and 8 ft/sec. The system is designed to accommodate all loads and load changes without adjusting these slurry circulation rates:

The following loops use this constant flow principle:

- Scrubber mill or first stage slurry recycle loop
- Second stage slurry feed
- Thickener feed
- Thickener underflow (sludge disposal)

Pressure Drop Pressure drop in either venturi stage is manually adjusted by raising or lowering the weir in a level tank outside the scrubber. Each stage can be adjusted independently to produce a pressure drop from 5 to 15 in. W.C. Raising or lowering the weir causes the slurry level in the scrubber near the lower edge of the venturi to rise or fall. The venturi pressure drop is linearly related to the slurry level in the vicinity of the venturi so that weir adjustments produce proportional changes in pressure drop.

The pressure drop, once set, is virtually unaffected by changes in gas flow rate. This insensitivity of pressure drop to gas flow results from the self-adjusting action of the slurry level in each venturi. When an increase in gas flow tends to increase the pressure drop, the slurry level tends to drop because of increased pickup of slurry by the gas stream. The drop in slurry level causes a decrease in pressure drop. When gas flow rate decreases, the slurry level rises increasing the pressure drop. Thus, the pressure drop is self-compensating as the gas flow varies and tends to stabilize at a value near the initial setting.

Automatic Controls

There are three essential automatic controls in the R-C/Bahco scrubber: reagent feed, slurry density, and makeup water or system level.

Reagent Feed The reagent feed system at RAFB is designed to maintain a preselected reagent-SO₂ stoichiometry for any load condition and any coal sulfur content within prescribed limits. Both the gas rate and the SO₂ concentration to the scrubber are measured continuously. These measurements are combined in a ratio controller which can regulate the reagent feed rate over a range of 20 to 1 to maintain the desired reagent-SO₂ stoichiometry.

Slurry Density Controlling the variations in slurry density is important for proper operation of the sludge de-watering system and to minimize scale formation. The slurry density control system at RAFB operates between set points of 10 and 12% solids. A sensor monitors slurry density and a controller is activated to allow thickened sludge at 40% solids to flow to the pond when the density reaches approximately 12% solids. Sludge flows continuously to the pond until the density in the system drops to 10% solids. When this point is reached, sludge is recycled to the scrubber and the line to the pond is flushed with water. This switching process is repeated as necessary to maintain slurry density in the desired range.

Makeup Water The total water requirement for the system varies almost directly with load since evaporative cooling of the flue gas consumes the bulk of the water used. Water is added to the system at several locations including the lime slaker and slurry pump seals. The balance of the makeup water is added through six spray manifolds located inside the scrubber module. The amount of water added through these sprays is regulated by a level sensor located in the lime dissolving tank. This sensor activates a programmed controller which adds water in a preselected sequence. Water losses from evaporation and sludge removal cause the level in the dissolver to drop initiating the water addition cycle to maintain the desired system operating level.

SYSTEM OPERATION

Start-up

The March, 1976 start-up of the the R-C/Bahco scrubber system proved to be a straightforward, one-day activity as a result of a careful, detailed check-out phase. The major equipment was checked out and operated with air and water for a few days prior to start-up. Operation on manual control for a few weeks after start-up proved simple and required very little attention. Switching the three basic control loops (lime feed, slurry density and tank level) over to automatic was accomplished smoothly and with a minimum of effort.

Research-Cottrell maintained around-the-clock shift coverage for one week after start-up. After the first week of operation, RAFB Heating Plant personnel took over operation of the scrubber. R-C continued to provide one engineer on stand-by for an additional month.

Problems Encountered

The problems encountered during the early operation of the R-C/Bahco system were mechanical and operational rather than process related. The scrubber operation itself has been excellent since start-up. No outages have occurred as a result of scrubber problems such as plugging or scaling. High lime reagent utilization and continuous slurry particle size reduction in the scrubber mill section appear to have contributed significantly to trouble-free scrubber operation.

System downtime has been related almost completely to auxiliary equipment failures and has occurred in two major areas. First, problems centered on the booster fan resulted in several periods of downtime for diagnosis and repair, especially during the early months of operation. Fan-related problems, particularly excessive vibration caused by bearing and support difficulties, were by far the major contributors to downtime.

The second category consisted of problems in a number of different areas. A thickener rake fabrication error discovered when the assembly was being installed contributed to downtime. Minor difficulties caused by pump, lime slaker and instrumentation installation errors or manufacturing defects also were encountered. Each of the problems has been solved and the system continues to operate very well.

The problems contributing to downtime occurred, for the most part, during the first months after start-up when the last few equipment and control items were installed and shakedown completed. Except for a several week shutdown caused by a fan malfunction, the R-C/Bahco scrubber availability has been considerably greater than 90% since the Air Force took beneficial occupancy in September, 1976. Availability has been greater than 98% since April, 1978.

Scrubber Inspections

During the first fifteen months of operation, a number of inspections were made of the scrubber internals when downtime permitted. The inspections were used to monitor the effectiveness of the water makeup system to keep areas clean that were potential sources of solids accumulation. The

initial inspection, a month after start-up, revealed an accumulation of solids in the second stage venturi slurry pan. It posed the only problem to smooth scrubber operation. The pan was emptied, solids were dislodged from the wet/dry zone and removed from the system. Subsequent investigation indicated that the lime slaker grit removal circuit in the unslaked lime was entering the scrubber and accumulating in the second stage slurry pan. The grit removal circuit was adjusted to remove this material and eliminate the problem.

Two additional precautions were found, as a result of the inspections, to be effective in improving scrubber operation. Increasing the blow-down frequency prevented accumulation of solids in the slurry outlets. Also, the slow accumulation of solids in the slurry outlets. Also, the slow accumulation of solids in the stack and gas straightening vanes at the top of the scrubber was further reduced by operating the second stage slurry at a lower pressure drop. In this way, the possibility of slurry droplet carryover was minimized.

No significant problems related to the accumulation of solids were found in the R-C/Bahco scrubber. The scrubber has the ability to tolerate substantial solids accumulation before scrubbing performance is adversely affected. Deterioration in performance, if it occurs, is gradual and usually can be rectified conveniently during a scheduled shutdown.

SO₂ REMOVAL PERFORMANCE

The R-C/Bahco scrubber is a high performance SO₂ removal device. The two-stage venturi unit at Rickenbacker very effectively scrubs SO₂ to low levels using a lime or limestone reagent.

Lime Reagent

Figure 4 shows that the SO₂ emission rate from the scrubber using lime is much lower than either the R-C/Bahco contract guarantee emissions rate of 1.0 lbs. SO₂/MM BTU (at 1.4 stoichiometry) or the state of Ohio limit for Rickenbacker Air Force Base of 2.2 lbs. SO₂/MM BTU. A stoichiometry of only 1.1 moles lime/moles SO₂ was required at Rickenbacker to almost completely eliminate SO₂ from the scrubber stack gas.

Of course, complete SO₂ removal is not required by the state of Ohio but stringent SO₂ emissions levels have been set for certain other sites. For example, emission rates as low as 0.3-0.4 lbs. SO₂/MM BTU have been promulgated for Wright Patterson AFB boilers at Dayton, Ohio. As Figure 4 shows,

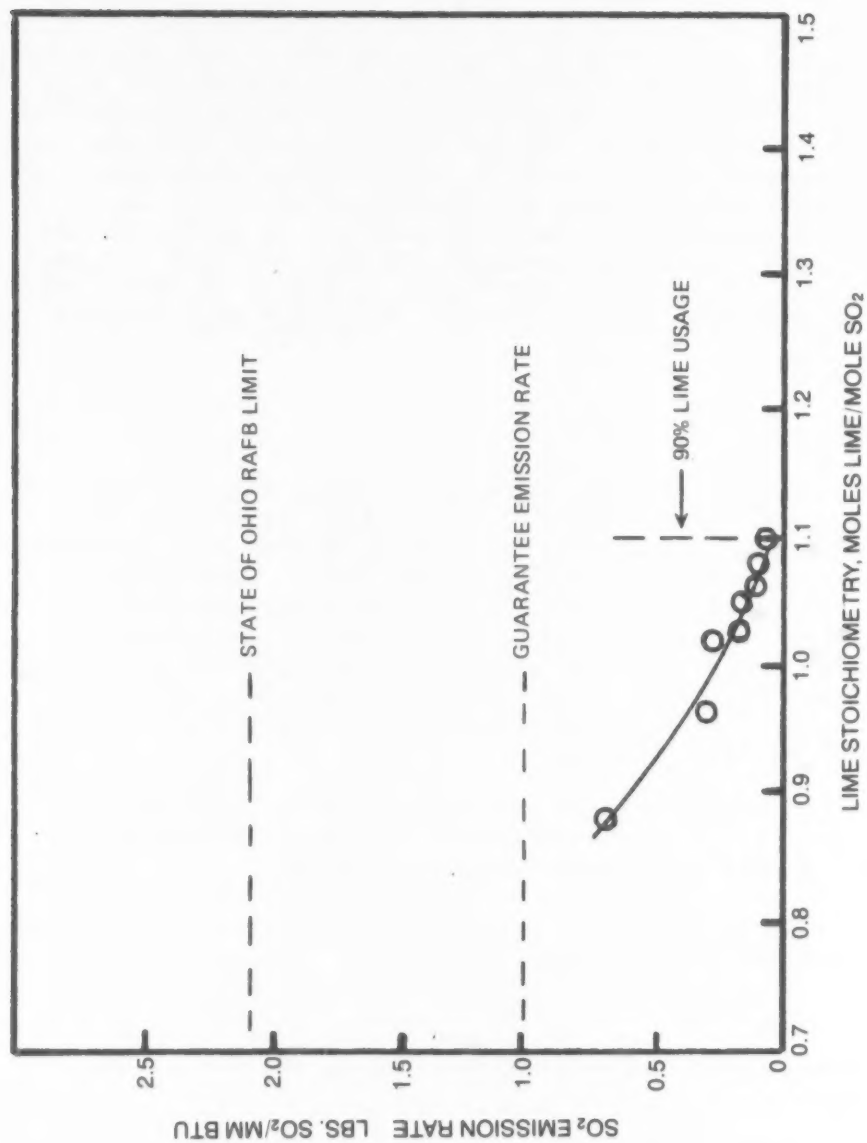


FIGURE 4: The relationship between SO₂ emission rates and lime stoichiometry.

these low SO₂ emission rates were attained at Rickenbacker using a lime/SO₂ stoichiometry of 1.0 or less when the boiler was burning 2.5-3.5% sulfur coal. The R-C/Bahco scrubber experience at RAFB indicates that, if necessary, the unit can be designed to remove SO₂ to easily comply with even the most stringent emissions requirements.

Over the range of conditions studied, the R-C/Bahco scrubber SO₂ removal efficiency was found to be a function only of the lime/SO₂ stoichiometric ratio. As Figure 5 shows SO₂ removal efficiency approaches 100% as the lime/SO₂ stoichiometry exceeds 1.0. These results were obtained with inlet SO₂ concentrations in the range of 500-2000 ppm and a total system pressure drop of 15-20 inches W.C. In this pressure drop range, the degree of mixing is sufficiently great to permit rapid reaction between the SO₂ and lime. Lime utilizations in the range of 90-100% were achieved in all tests performed at Rickenbacker even those at very high SO₂ removal efficiencies.

Limestone Reagent

Limestone reagent can also be used to meet the requirements for SO₂ removal at industrial coal-fired installations. Tests at RAFB indicated that SO₂ removal of 80-90% was achieved at limestone stoichiometries of 1.0 or greater. Limestone stoichiometry and slurry feed rate to the scrubber significantly affect SO₂ removal, as Figure 6 shows. Increased stoichiometry and slurry feed rate result in higher SO₂ removal. Limestone utilizations range from about 60% at conditions where higher SO₂ removal percentages are achieved to 90% at lower SO₂ removal values.

The effect of stoichiometry and slurry feed rate on SO₂ removal in the R-C/Bahco scrubber using a limestone reagent were correlated by the following empirical model:

$$\% \text{ SO}_2 \text{ Removal} = (\text{St})^{0.52} (\text{L})^{0.55} \quad (1)$$

where:

St = Stoichiometry, moles CaCO₃/mole SO₂ in the inlet gas and,

L = Slurry feed rate to the second stage, gpm.

Figure 7 shows a comparison between the SO₂ removal predicted by Equation (1) and observed SO₂ removal. Actual SO₂ removal performance and the model prediction fall within $\pm 15\%$ of each other indicating that a satisfactory relationship was obtained from the field tests.

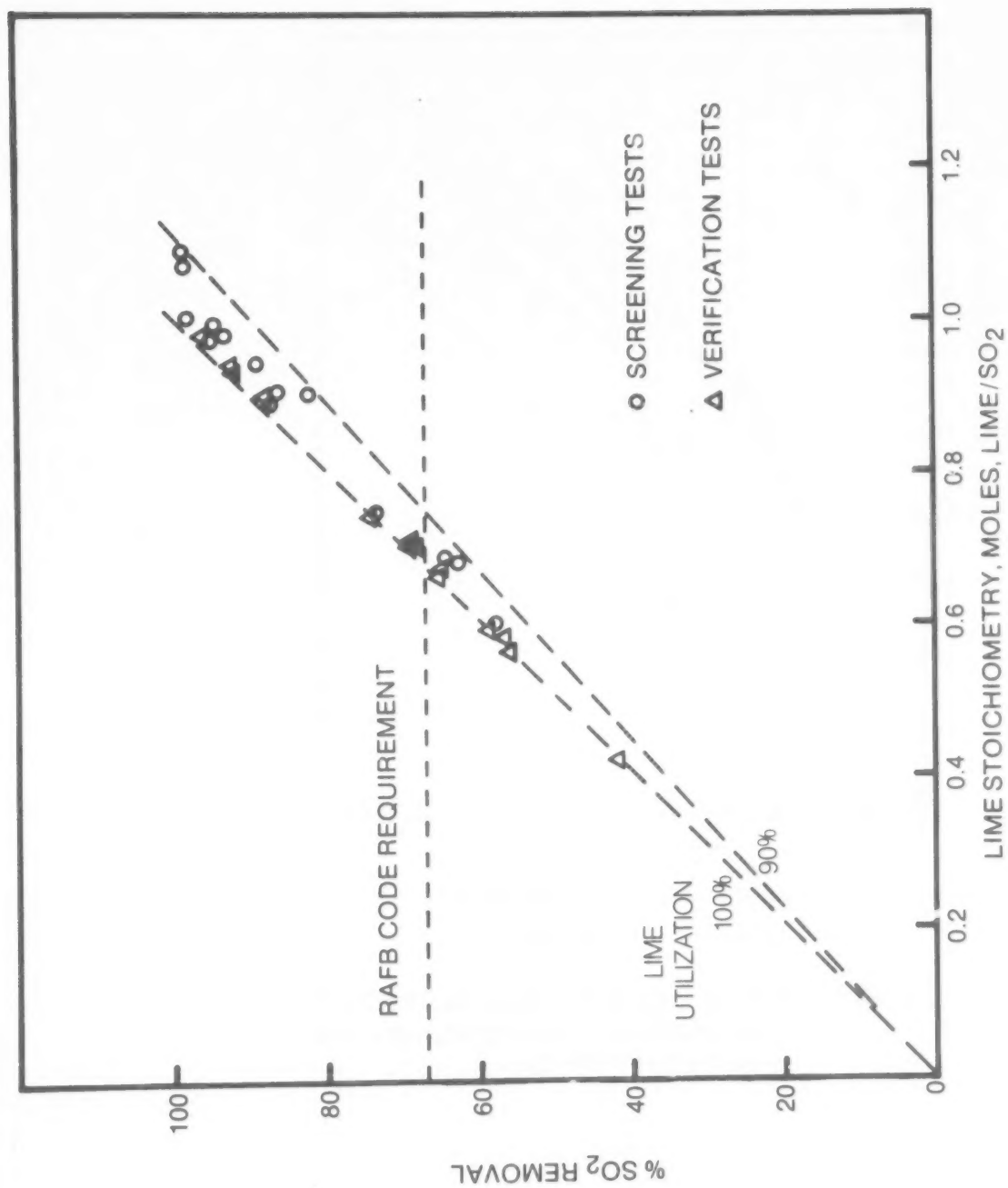


FIGURE 5 SO₂ removal efficiency as a function of lime stoichiometry.

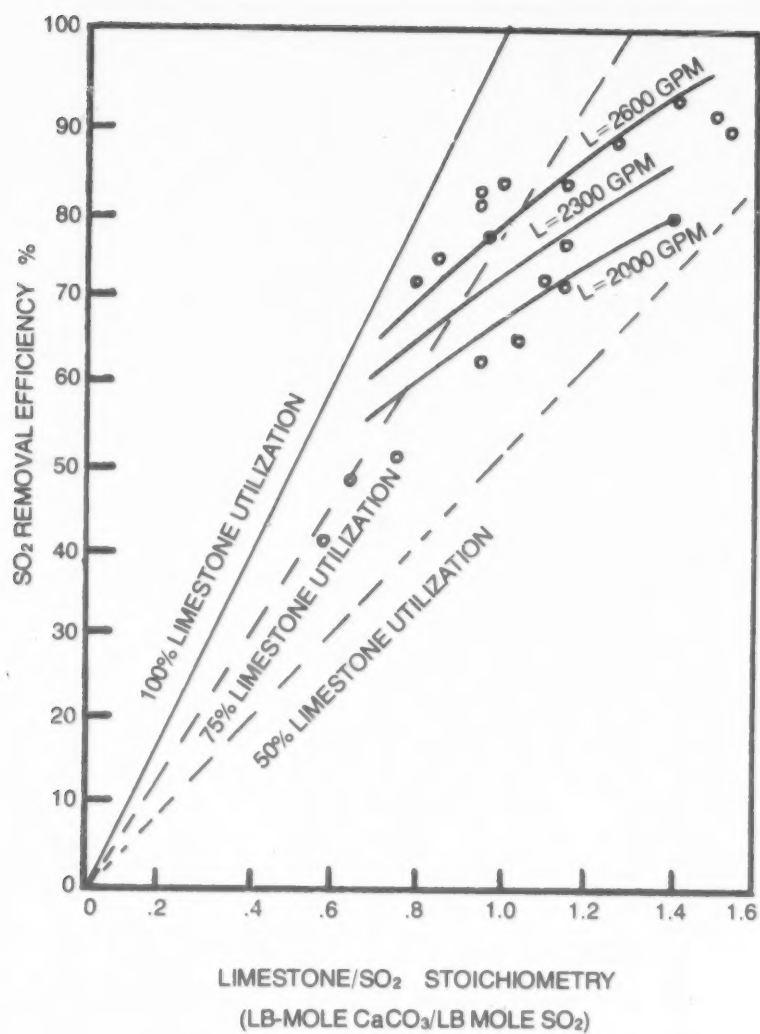


FIGURE 6: SO₂ removal efficiency as a function of limestone/SO₂ stoichiometry and slurry pumping rate.

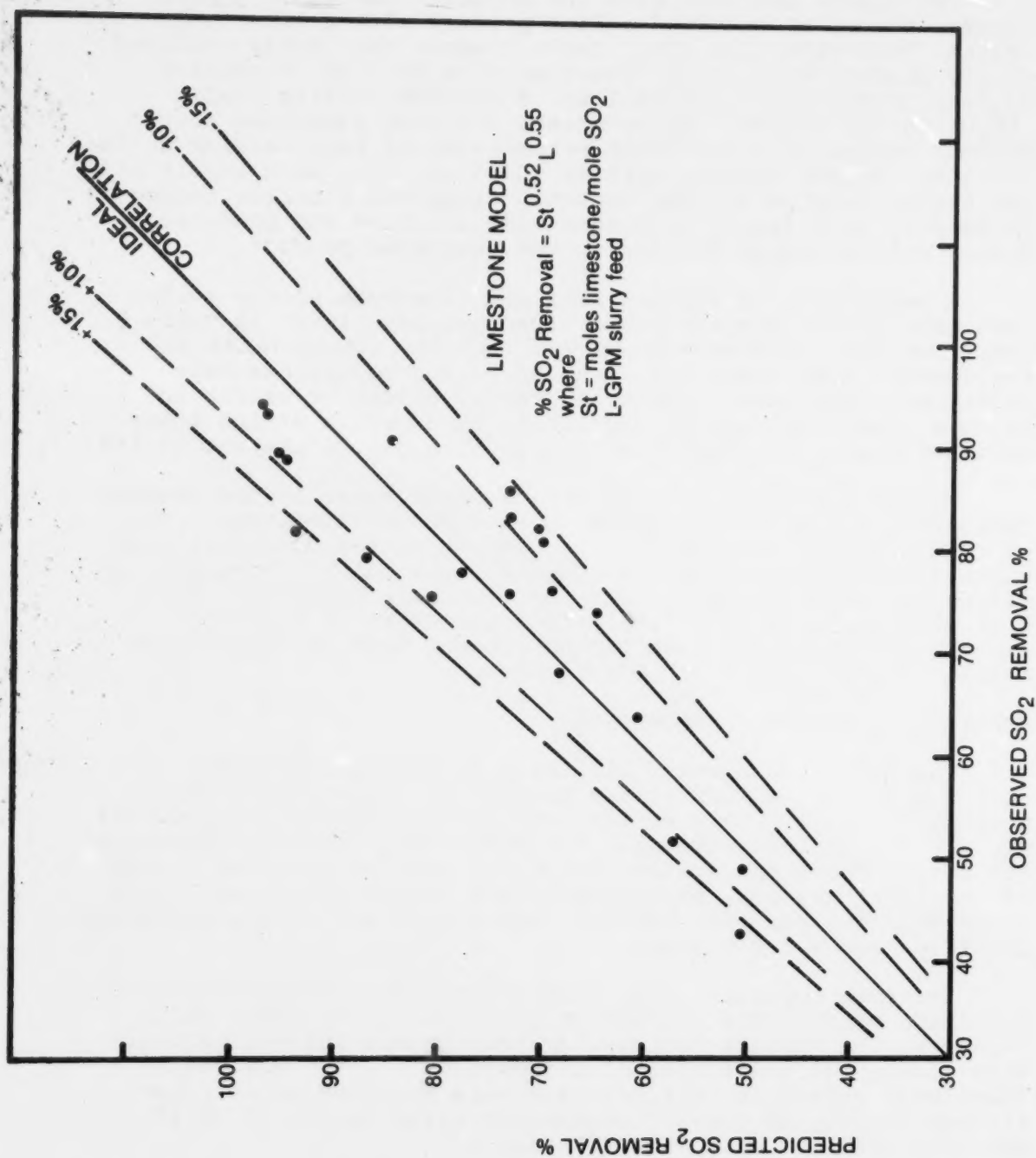


FIGURE 7: A comparison of predicted and observed SO_2 removal efficiencies for limestone.

Scrubber Slurry Product

The slurry effluent from the scrubber contained a much greater amount of oxidized sulfur product with a limestone reagent than with lime (3). Table 1 shows the highly oxidized slurry product when using limestone with 62.3 wt. % calcium sulfate ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$) and 16.1 wt. % calcium sulfite ($\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$) in the solids. In contrast, the lime generated slurry product contained appreciable percentages of both calcium sulfate (33.4 wt. %) and calcium sulfite (54.5 wt. %). As a result of the higher calcium sulfate content, limestone slurries tended to dewater at a faster rate than lime slurries and produced higher solids concentrations in the dewatered product.

A comparison of average lime and limestone slurry analyses made during similar boiler load periods listed in Table 1 indicates the oxidation trend was directly attributable to the reagent type since all other operating conditions were essentially the same. The much greater extent of oxidation using a limestone reagent may likely be a result of the lower scrubber slurry pH coupled with high flue gas O_2 content ($\geq 14\%$).

Figure 8 shows slurry pH measurements taken in the reagent dissolver tank during operation with lime and limestone. The pH of the slurry recycle loop is highest in the dissolver tank since fresh basic reagent is added at this point. Dissolver pH values are significantly lower for limestone operation (4.9-6.2) than for lime (4.3-9.6). The high level of oxidation experienced with limestone in the present work contrasts with results obtained in other FGD studies.

PARTICULATE REMOVAL PERFORMANCE

The R-C/Bahco system utilizes a mechanical collector in series with two venturi stages located in the scrubber to remove particulate. This series combination results in up to 99% fly ash particulate removal. The mechanical collector protects the forced draft fan against excessive wear by removing 70-80% of the total fly ash, particularly the larger particles (>5 -10 microns). The scrubber removes essentially all of the remaining particles above 1-2 microns.

Andersen impactor tests were conducted at Rickenbacker to determine particulate collection efficiency. As shown in Figure 9, the results indicate an aerodynamic cut-off particle size (i.e., 50% collection efficiency) of 0.6-0.7 microns. These high collection efficiencies were obtained at an inlet fly ash loading of 0.2-0.3 grains/SCF using an L/G of 15-25 gpm/1000 ACFM for each venturi stage.

TABLE 1
LIME AND LIMESTONE SLURRY ANALYSES

Slurry Solids	Lime Slurry Limestone Slurry	
	Weight %	Weight %
$\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$	33.4	62.3
$\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$	54.5	16.1
CaCO_3	3.7	14.2
MgCO_3	—	0.8
Acid Insolubles & Others	8.4	6.6
TOTAL	100.0	100.0

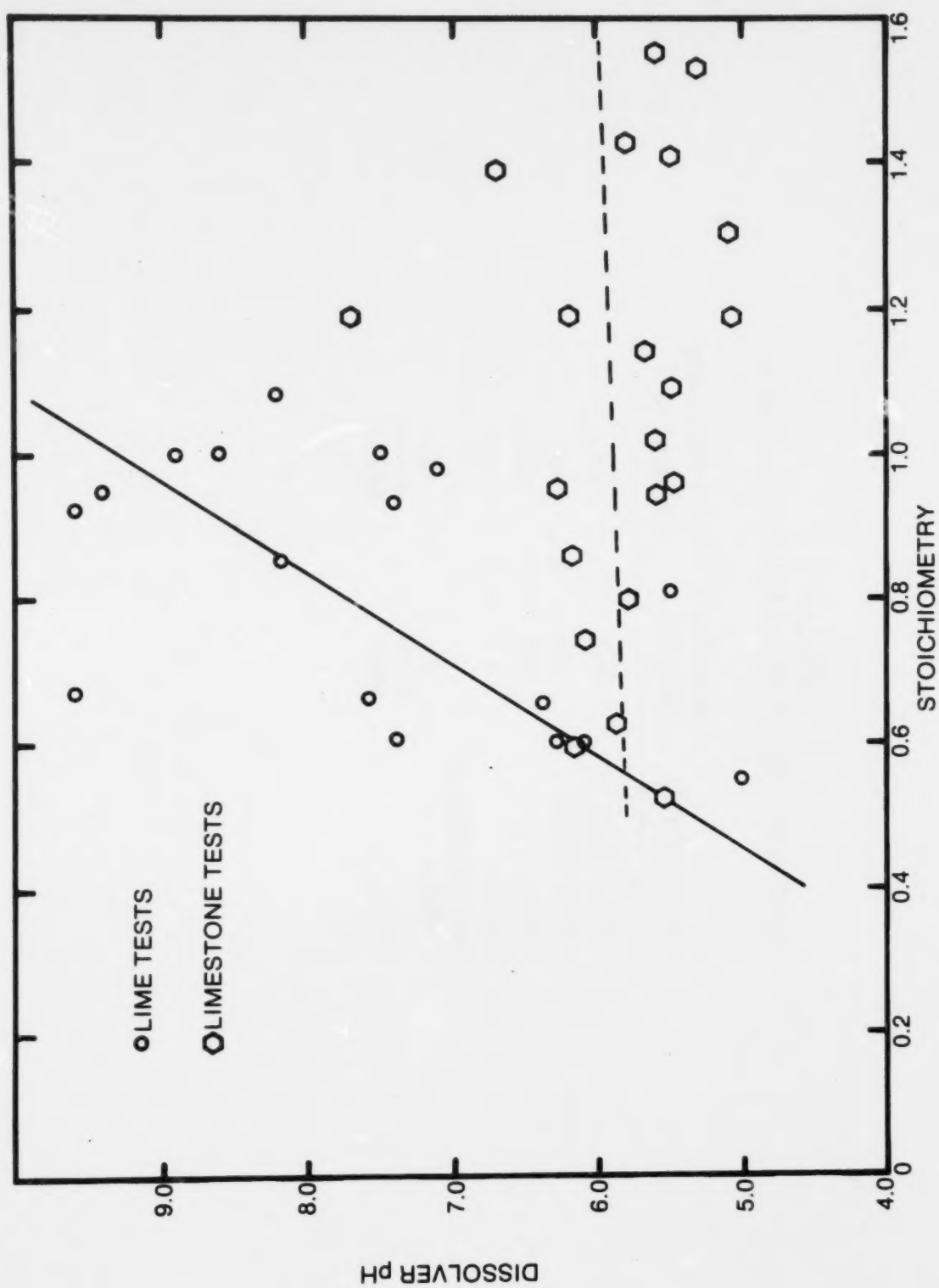


FIGURE 8: The effect of lime and limestone stoichiometry on dissolver pH.

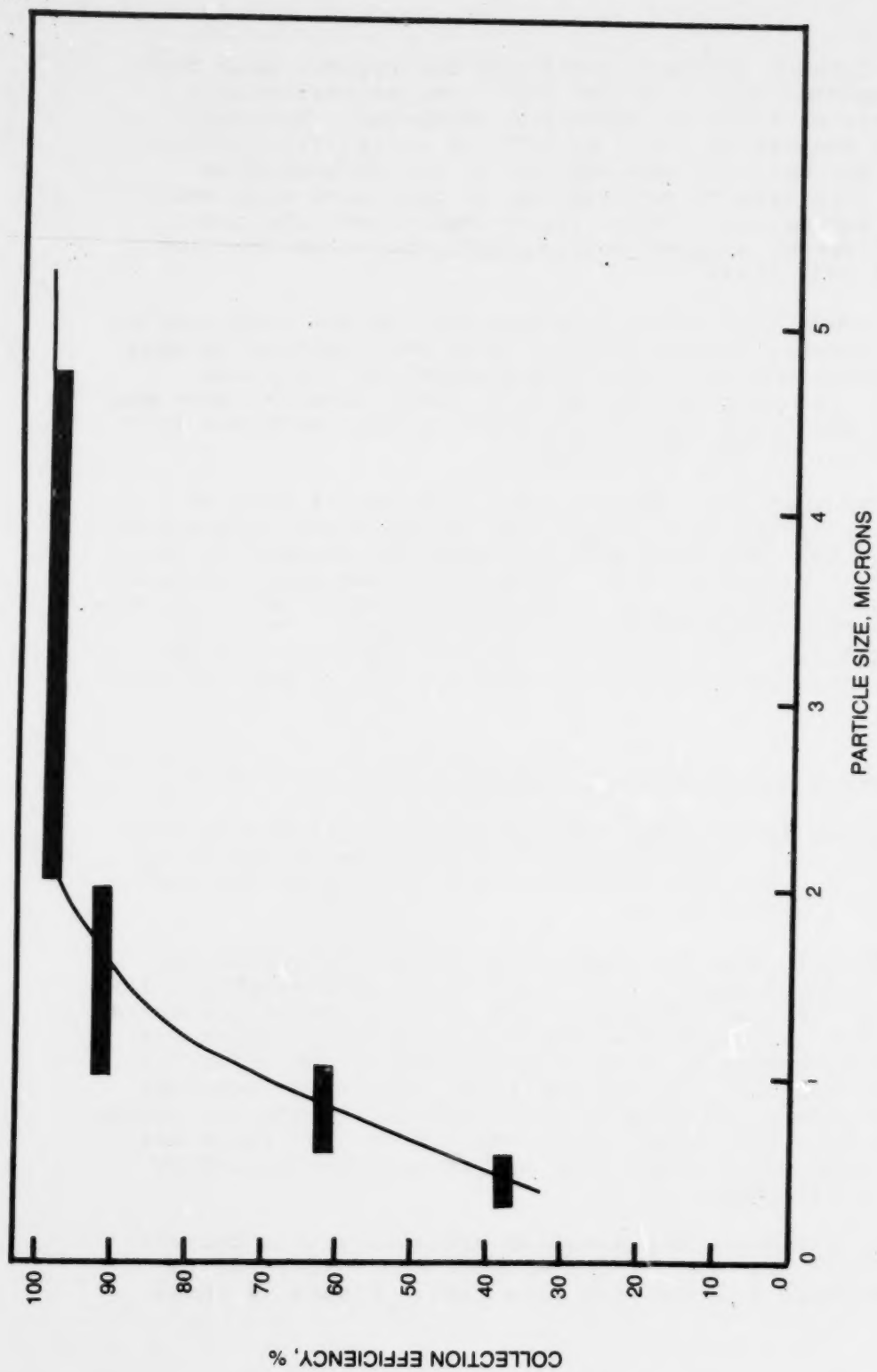


FIGURE 9: R-C/Bahco System Collection Efficiency.

The R-C/Bahco scrubber exhibited the typical high performance expected from a device that removes particulate primarily via an inertial impaction mechanism. However, significant amounts (0.1-0.2 gr/SCF) of sooty, fine particulate (<1 micron) were encountered in the Rickenbacker flue gas. This type of particulate is generated when combustion is incomplete. Under these conditions, the collection efficiency dropped significantly for this fraction, as Figure 9 indicates.

Under conditions where fine particulate was generated at RAFB, considerably higher pressure drop was required to meet guarantee emissions rate than anticipated for a fly ash particulate. At least 18 inches W.C. total pressure drop was employed in the two R-C/Bahco scrubber venturi sections to meet guarantee, as Figure 10 shows.

The amount of soot found in the flue gas at RAFB is substantially higher than normal for stoker-fired generators. As a result, the USAF undertook an extensive program to upgrade the Heat Plant at RAFB. Information Research-Cottrell obtained during the EPA test program contributed significantly to the specific tasks of the upgrading program. This incident illustrates how pollution control findings may be translated into substantial fuel cost savings in the boiler operation.

R-C/Bahco Fractional Efficiency Results

Fractional particulate removal efficiencies by the R-C/Bahco scrubber were obtained over a wide range of operating conditions. Key process variables were related to 90% and 50% collection efficiencies.

The particle cut-off diameter at which 90% collection efficiency was obtained varied from 0.7 to 1.25 microns. Increasing the total scrubber pressure drop increased collection efficiency and decreased the particle diameter at which 90% collection was observed. Figure 11 indicates the effect of the combined pressure drop of the first and second scrubber stages on the particle size at which 90% collection was achieved. Particulate collection efficiency in the 90% range was only a function of pressure drop and appears unaffected by other system variables.

For 50% particulate collection efficiency, the cut-off diameter ranged from 0.4 to 0.7 microns and is related to scrubber pressure drop and gas flow rate. Figure 12 shows

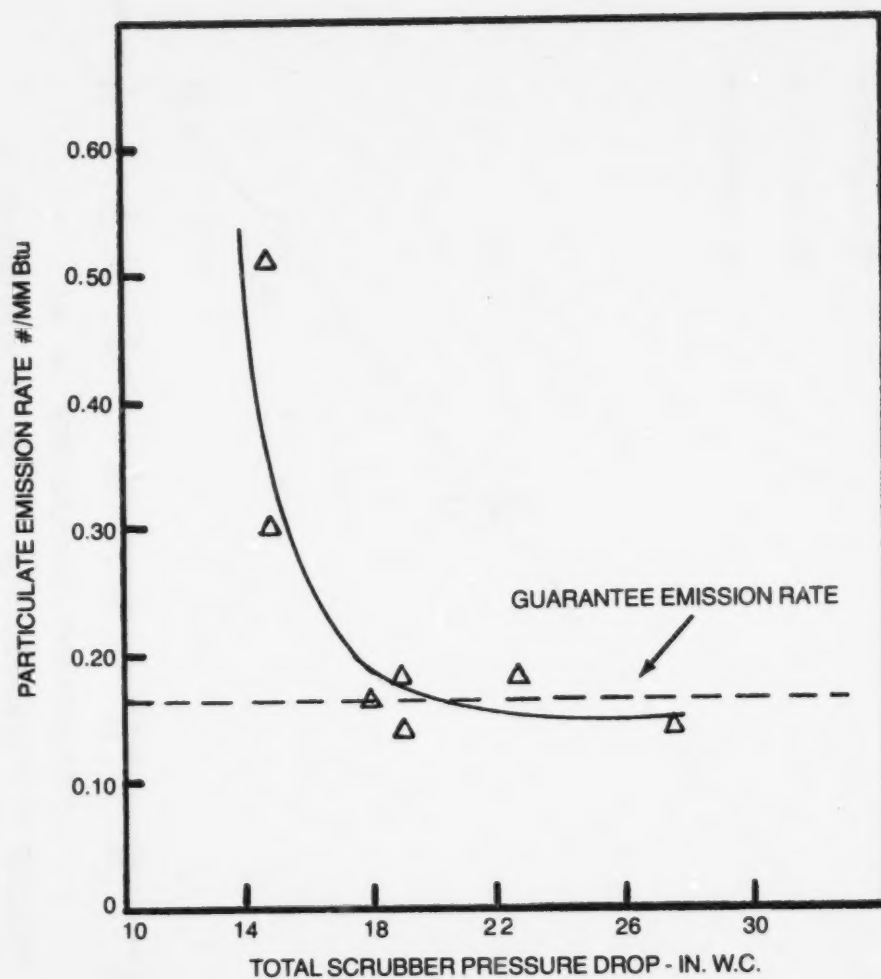


FIGURE 10: The effect of scrubber pressure drop on particulate emission rates.

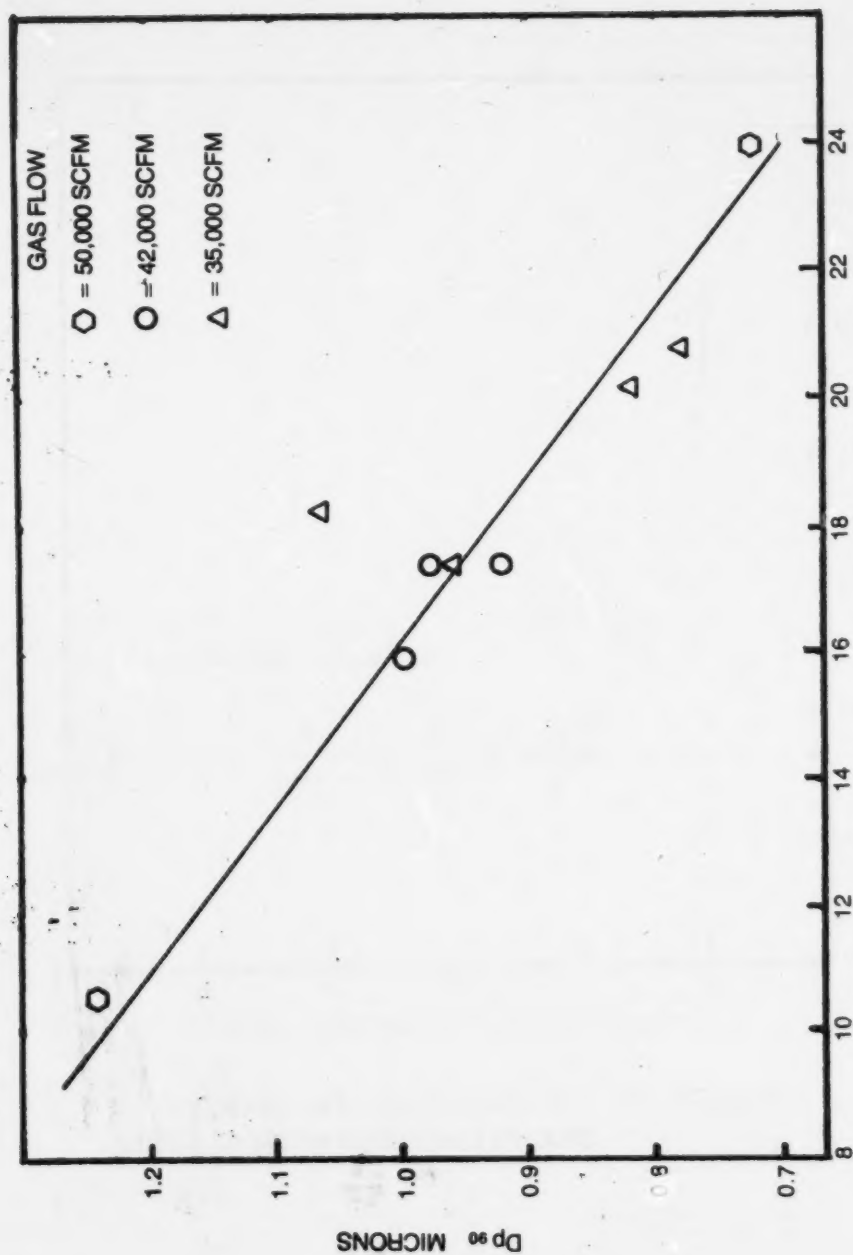


FIGURE 11: The effect of total scrubber pressure drop on the cut off diameter for 90% collection efficiency.

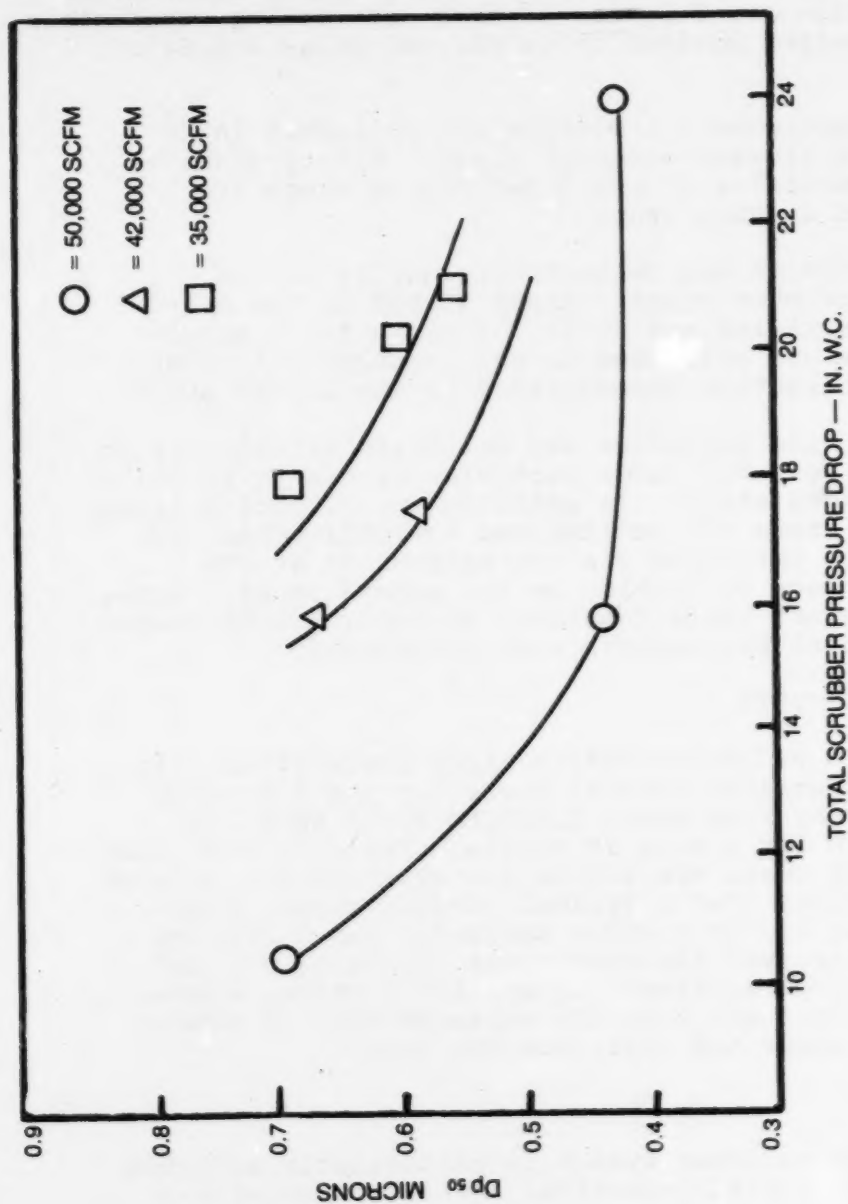


FIGURE 12: The effect of total scrubber pressure drop and gas rate on the cut off diameter for 50% collection efficiency.

that increasing the pressure drop and the gas rate enhances particulate collection as indicated by the decrease in the diameter observed for 50% collection efficiency.

Further analysis of the fractional efficiency information using a penetration model led to several interesting conclusions regarding particulate collection in the two stage R-C/Bahco scrubber:

- Larger particles (>1 micron) are collected in the first (lower) scrubber stage. Ninety-eight % of the particles in the 2.0-5.0 size range are collected in this stage.
- Fine particles are collected primarily in the second scrubber stage. About 40-50% of the 0.3-0.5 micron particles and 60-65% of the 0.5-1.0 micron particles are collected in the scrubber with most of the collection accomplished in the second stage.

Since large size particles are collected efficiently in the first stage, very few large particles remain to be collected in the second stage. In addition to collecting large particles, first stage gas cooling and humidification may condition the fine particles via nucleation and growth mechanisms to enhance collection in the second stage. Thus, each scrubber venturi stage functions to collect a different portion of the total particulate size distribution.

R-C/BAHCO CAPITAL COSTS

Nine standard R-C/Bahco module sizes are available for gas treatment. Installed capital costs for the R-C/Bahco scrubber system vary from about \$35/SCFM for a size 100 module to \$50/SCFM for a size 20 module. Based on 2000 SCFM/MW, the equivalent costs are \$70/KW for size 100 and \$100/KW for a size 20 module. For a typical project scope, Figure 13 presents R-C/Bahco system turnkey costs for the different size modules. Site specific conditions including SO₂ and particulate cleaning requirements, available space, sludge disposal requirements and ductwork arrangements, of course, will dictate the scope and influence the cost.

Size Selection

The R-C/Bahco scrubber system is particularly suitable for small and medium size industrial installations with a single module treating a gas stream in the range of 10,000 to 90,000 SCFM (4). This range is roughly equivalent to a

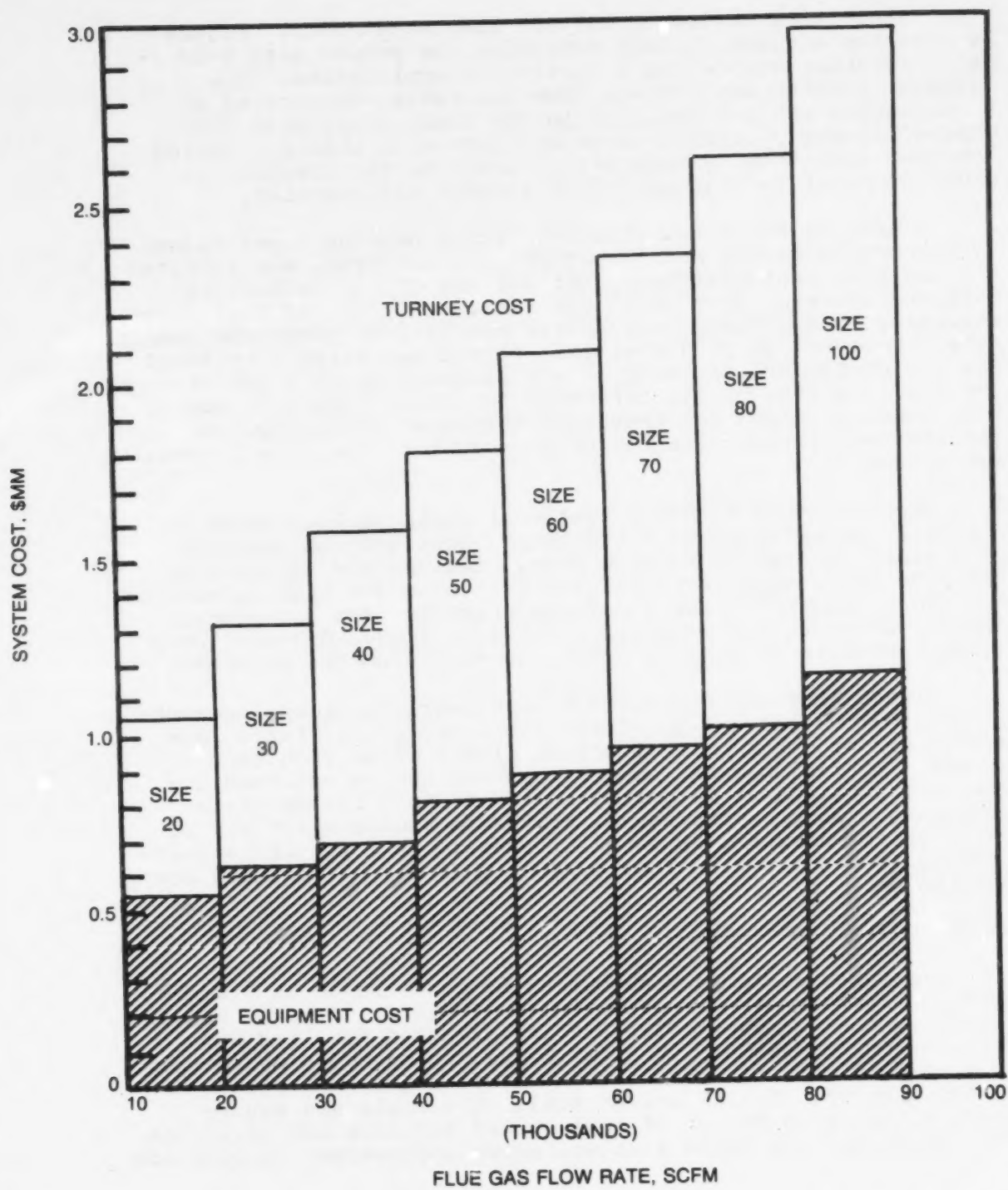


FIGURE 13: R-C/Bahco Capital Costs.

plant generating 30,000 to 300,000 lbs. steam/hour. Figure 14 presents a chart to help determine the proper size R-C/Bahco scrubber module for a particular application. The maximum, minimum and average flue gas rates encountered at Rickenbacker AFB are depicted in the chart along with the most efficient operating range for each size module. During low load operation, makeup air is added to the scrubber in order to maintain high pollutant removal efficiencies.

A size 50 R-C/Bahco scrubber, which handles a gas volume corresponding to 200 million BTU/hr. firing rate, was selected for the RAFB installation. Combinations of the boilers at RAFB are operated from 58 to 208 million BTU/hr. of fuel fired, resulting in a maximum winter rate nearly four times the summer rate. However, in 1973 when the project was first considered the 200 MM BTU/hr. firing rate was exceeded on only one day (by 5%). In view of the rare need for extra capacity, the additional cost and the increased turn-down ratio required for the next larger size R-C/Bahco scrubber, the size 50 module was chosen.

The R-C/Bahco scrubber system at Rickenbacker, shown in Figure 1, occupies about 2,500 square feet and the scrubber plus stack is eighty-one feet high. Besides the scrubber, the lime silo, thickener, fan and control house take up most of the space required. The remaining space is used for pumps, piping, ductwork and access. A 2.5 acre sludge disposal pond at Rickenbacker is located about 750 feet from the scrubber.

Sludge handling development work carried out by Research-Cottrell since the design of the Rickenbacker Air Force Base system has produced more efficient sludge dewatering techniques. Hydroclones and centrifuges can now be utilized instead of a thickener to permit a more compact design while producing a more concentrated, truckable sludge which may be suitable for disposal in landfill areas. As a result, significantly less space than that used at RAFB may be required for the system.

ANNUAL OPERATING COSTS

R-C/Bahco system annual operating costs based on Rickenbacker AFB operation are presented in Table 2. A total annual operating cost of about \$235,000 is estimated for lime and \$212,000 for limestone. The estimates are based on operation in a size 50 scrubber module treating flue gas generated by 40,000 tons/yr. of 3.5% sulfur coal. The costs are equivalent to about \$5.90/ton of coal burned for lime and \$5.30/ton for limestone. As Table 2 shows, power and reagent consumption

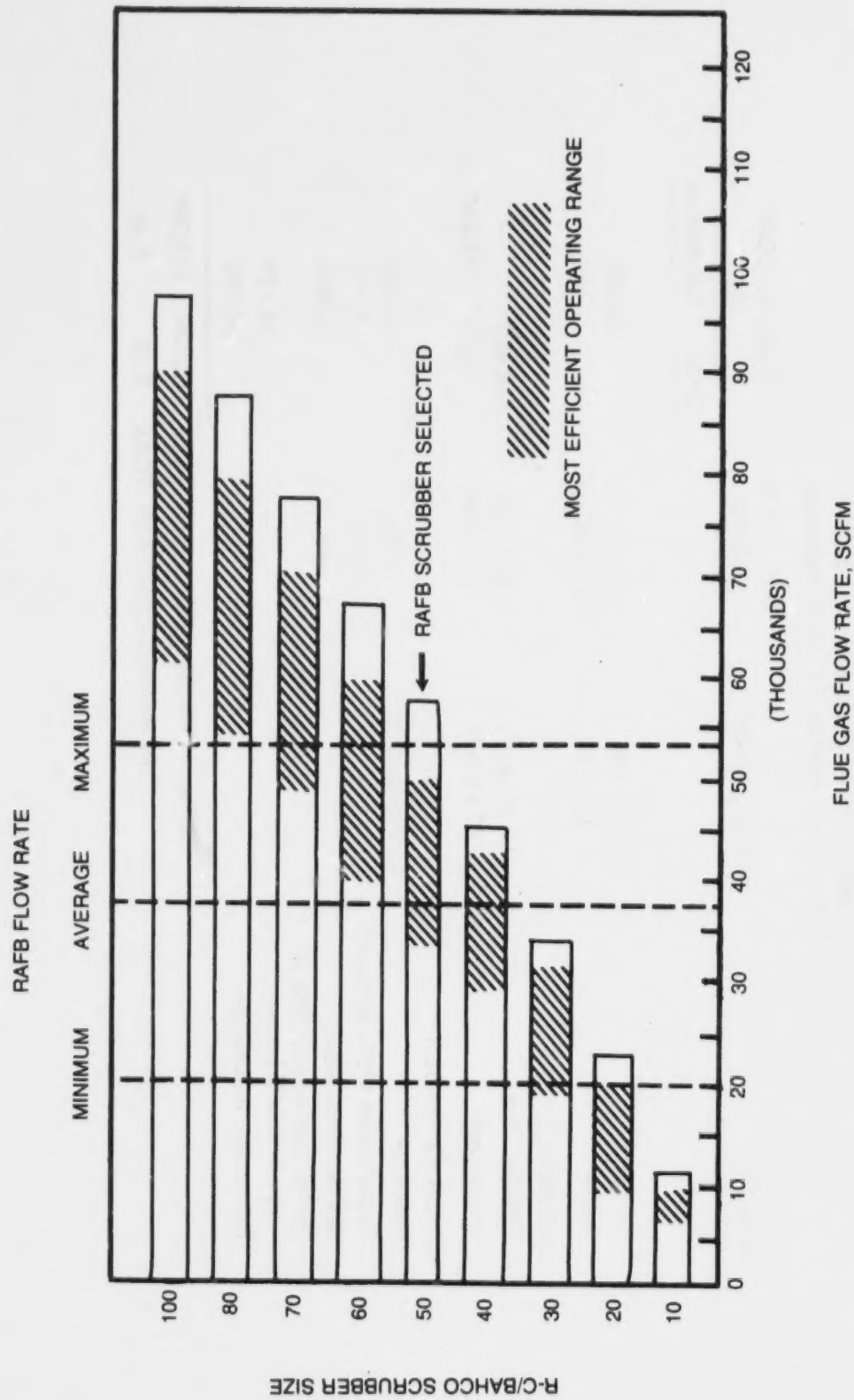


FIGURE 14: R-C/Bahco Scrubber Sizing.

TABLE 2
R-C/BAHCO ANNUAL OPERATING COSTS
 (Based on 40,000 tons/yr. of 3.5%
 S coal consumption, 70% SO₂ removal)

<u>UTILITY</u>	<u>REQUIREMENT</u>	<u>UNIT COST (\$/Unit)</u>	<u>ANNUAL COST (\$)</u>	
			<u>Lime</u>	<u>Limestone</u>
Power	350 KW	0.027/KWH	78,250	
Water	10 GPM	0.37/1000 gal.	1,950	
CHEMICALS				
Pebble Lime	0.2 tons/hr.	40.35/ton	66,820	—
Limestone	0.41 tons/hr.	12.72/ton	—	43,390
OTHER EXPENSES				
Operator	0.25 man/shift	8/hr.	16,560	
Supervision (25% of labor)		10/hr.	5,170	
Maintenance M&L			12,600	
Direct Overhead (75% of op. & Maint. Labor)			19,170	
Taxes and Insurance			35,000	
TOTAL ANNUAL OPERATING COST			235,520	212,090
OPERATING COST/TON COAL			5.90	5.30

are the two principal contributors to expenses, comprising more than half of the annual operating cost. During periods of inefficient boiler operation at Rickenbacker, where very high excess air (120-250%) was utilized or copious amounts of soot were generated, up to 35% additional power was employed to cope with the situation.

At RAFB, limestone is more economical to use than lime despite the fact that more than twice as much limestone is needed to obtain the same SO₂ removal. The price of limestone delivered to RAFB during 1977 was \$12.72/ton compared to \$40.35/ton for lime. This large price differential means that for every \$100 spent for limestone, greater than \$140 must be spent for lime for equivalent performance to meet code.

Limestone has other advantages over lime because it is not hygroscopic and need not be slaked thereby eliminating the need for a slaker. Also, limestone is less likely to cause injuries to operating or maintenance personnel since it does not exhibit the caustic properties inherent to lime. For these reasons, RAFB switched from lime to limestone reagent in May, 1977, and has since experienced smooth, effective scrubber operation as well as considerable cost savings.

The R-C/Bahco system requires only a minimum amount of operator attention. At RAFB, a scrubber technician operates the system on day shift Monday through Friday and also handles routine maintenance. Rickenbacker Heat Plant operating personnel monitor the scrubber operation during off shifts and on weekends. The cost for operating labor, including supervision and overhead, plus maintenance material and labor is about \$55,000/yr.

Sludge disposal operating costs are minimal at RAFB since the sludge is pumped to a disposal pond. The installed pond cost amounts to \$0.45/ton of coal.

ACKNOWLEDGEMENTS

We wish to thank Mr. James Rasor, Base Civil Engineer, Rickenbacker Air Force Base, Lockbourne, Ohio for his help and cooperation in enabling Research-Cottrell to conduct the R-C/Bahco test program at Rickenbacker. The direction and aid of Mr. John Williams, EPA Project Officer, is also greatly appreciated. R-C personnel who contributed heavily to the commercial job and the EPA test program are Robert Ferb - Project Manager, Edward Biedell, David Ruff and Gary Malamud.

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Status of the Project to Develop Standards of Performance for
Industrial Fossil-Fuel-Fired Boilers

L. D. Broz, G. R. Offen, P. D. Anderson,
Acurex Corporation
Raleigh, NC

and

J. D. Mobley, C. B. Sedman,
U.S. Environmental Protection Agency
Research Triangle Park, NC

ABSTRACT

The Environmental Protection Agency has undertaken a study of industrial boilers with the intent of proposing standards of performance based on information gathered during the project. The study is being directed by EPA's Office of Air Quality Planning and Standards, and technical support is being provided by the Agency's Industrial Environmental Research Laboratory at Research Triangle Park, N.C. Acurex is the systems integration contractor for both groups on this project. Other EPA offices and support contractors are also involved.

Through a series of tasks, background information is being collected on industrial coal-, oil-, and gas-fired boilers and the technologies that have demonstrated the ability to reduce nitrogen oxides, sulfur oxides, and particulate emissions from these sources. This information collection activity is nearing completion and will be followed by extensive analyses of the potential economic, energy, and environmental impacts of alternative regulatory options. The data and analyses will be documented in a Background Information Document and used to develop and support a New Source Performance Standard for the industrial boiler source category.

INTRODUCTION

The Clean Air Act, as amended in 1977, provides authority for the U.S. Environmental Protection Agency (EPA) to control the discharge of air pollutants into the atmosphere. The Act contains several regulatory and enforcement options for control of emissions from stationary sources. Options include (1) National Ambient Air Quality Standards (NAAQS) on the national level and State Implementation Plans (SIP's) on the state level, (2) new source performance standards (NSPS) on the federal level, and (3) national emission standards for hazardous air pollutants (NESHAPS).

Section 111 of the Act calls for issuance of emission standards for emissions from new and modified sources which may contribute significantly to air pollution, and which could endanger public health or welfare. The standards must reflect the best degree of control, taking cost, energy, and nonair environmental quality impacts into account. No new plant can be built, however, if it will violate a NAAQS, even if it meets the NSPS.

Amendments to the Act in 1977 specifically mention the need to develop standards for fossil-fuel-fired boilers. Further, in a prioritized list of sources for which standards of performance should be developed, industrial boilers rate 13th of 66 source categories. Accordingly, EPA has undertaken a study of industrial boilers with the intent to propose standards of performance, or NSPS, based on information gathered during the project. Further, to accommodate the accelerated timetable for the development of NSPS, as mandated by the 1977 amendments, EPA has greatly accelerated its use of contractors to develop these standards. Acurex Corporation, one of several contractors serving EPA in this capacity, has been assigned the industrial boilers source category.

Much of the background information concerning the control technologies for this source category has been or is being developed by EPA's Industrial Environmental Research Laboratory--Research Triangle Park (IERL-RTP) and the Office of Air Quality Planning and Standards (OAQPS). OAQPS is lead office on the project and ultimately responsible for development of the standard. In its support role to OAQPS, IERL-RTP is providing direct assistance by developing information concerning control technologies. OAQPS will use this and other information to prepare a Background Information Document (BID), which will form the basis for the standards to be proposed. The input from IERL-RTP will be in the form of Technology Assessment Reports (TAR's), which will be referenced in the BID. These reports will also document IERL-RTP's assessment of the state of the art in control technology for industrial boilers. In addition, they will be used within EPA's Office of Research and Development (ORD) to evaluate the proposed emission standards.

Acquisition of background information by IERL-RTP, OAQPS, and contractors is directed by OAQPS and will culminate in preparation of the BID. To avoid duplication of effort in preparing the TAR's and the BID, IERL-RTP and OAQPS are utilizing the same integration contractor, Acurex, directed by an OAQPS Task Manager.

Current Regulations for Industrial Boilers

One measure of the impact of an NSPS on emissions is an evaluation relative to typical state regulations. Hence, these limits are summarized briefly here. Most states restrict particulate matter (PM), sulfur oxides (SO_x) and nitrogen oxides (NO_x) emitted from industrial boilers. Typical emission limits range from 43 to 344 ng/J (0.1 to 0.8 lb/10⁶ Btu)

heat input for particulates and 86 to 3870 ng/J (0.2 to 9.0 lb/10⁶ Btu) heat input for SO_x. Typical NO_x emission limits range from 86 to 387 ng/J (0.2 to 0.9 lb/10⁶ Btu) heat input; however, 24 states have no regulations for NO_x emissions.

A federal standard of performance currently exists for large boilers. This standard applies to all fossil-fuel- and wood-fired steam generating units capable of firing at a heat input of more than 73 MW (250 x 10⁶ Btu per hour) and constructed or modified after August 17, 1971. EPA recently proposed a revision to this standard which will apply only to electric utility steam generating units. The existing NSPS for fossil-fuel-fired steam generators will continue to apply to industrial boilers with heat input greater than 73 MW (250 x 10⁶ Btu per hour), until revised as a result of this study.

Scope of Industrial Boiler Source Category

The industrial boiler source category currently under consideration for regulation by an NSPS includes all non-residential and non-utility boilers. Therefore, all institutional, commercial, and industrial boilers are included, although this study is commonly referred to as the industrial boiler NSPS activity. The population of industrial boilers consists of many different types of units, which use various fuels and methods of heat transfer. This study, however, deals only with fossil-fuel-fired boilers; separate EPA studies are evaluating the feasibility of standards on waste-fired boilers and incinerators. On the other hand, technologies such as fluidized bed combustors (FBC), synthetic-fuel-fired boilers, and fuel cleaning are being investigated.

Pollutants of Concern

Emphasis is being placed on the criteria pollutants emitted by combustion sources--PM, SO_x, and NO_x. Growth projections for industrial fuel usage vary over a wide range, but a value of 3.7 percent compounded annually appears reasonable. Based on this nominal growth rate projection, emissions from industrial boilers would increase to nearly 250 percent of their current level by the year 2000 if not controlled beyond present regulations.

Total particulate emissions depend primarily on the quantity of coal used and the ash content. Hence, as fuel conversion strategies are implemented, particulate levels will increase. Total SO_x emissions will change in direct relation to changes in fuel sulfur content of the fuels burned; thus, they will also increase somewhat beyond the levels that would be otherwise projected to the extent that coal replaces expected natural gas and distillate oil consumption. For the same reasons--growth and emphasis on coal utilization--NO_x emissions will change considerably from current levels (NO_x emissions from coal-fired units are 200 to 300 percent greater than from distillate oil or natural gas boilers).

GENERAL APPROACH

The general approach being taken to develop the NSPS for industrial boilers follows:

1. Characterize the source category and current emission rates, either controlled or uncontrolled as applicable, to meet SIP's.
2. Collect process information and performance data on boilers with emission control systems:
 - from available literature

- from emission tests at sites with candidate best demonstrated controls.
- 3. Identify potential changes to existing sources that could be deemed modifications or reconstructions, as defined by EPA regulations, and which thus would require compliance with the NSPS.
- 4. Select model plants (model boilers) for use in site specific economic, environmental and energy impact analysis. Model boilers are combinations of representative boilers and control systems.
- 5. Utilize a computer model to predict the potential economic, environmental, and energy impacts due to alternative regulatory options.
- 6. Determine the potential economic, environmental (all media), and energy impact due to each of the alternative regulatory options, using both the computerized analysis and model boilers.
- 7. Recommend a standard based on the best demonstrated technological system of continuous emissions reduction considering economic, environmental, and energy factors.

Formal proposal of the NSPS is preceded by circulation of a draft BID, which discusses the emission sources and emission control alternatives and assesses the performance, cost, energy requirements, and overall environmental impact of each alternative control system. Various environmental groups, industry, and other interested parties participate in formal review of the document which culminates in a public meeting before the National Air Pollution Control Techniques Advisory Committee (NAPCTAC). Following NAPCTAC review, the document and recommended standards are subjected to additional intragovernmental review before the final standards package is

approved for proposal by the Administrator. The proposed NSPS are then published in the Federal Register, inviting public comment. Once all comments have been received and resolved to the Administrator's satisfaction, final standards of performance are promulgated by publication in the Federal Register.

Project Organization

Figure 1 illustrates the organization of responsibilities. Basically it shows the following:

- The Emission Standards and Engineering Division (ESED) of OAQPS is the lead organization with ultimate responsibility for development of the standard. As such, it directs the identification and evaluation of regulatory options, considers public comments, and oversees the preparation of the documents needed to propose and promulgate a standard. In this study, ESED is:
 - Supported by Acurex, as systems contractor, to: (1) integrate the effort which leads to the BID and recommended standard; (2) provide support during the review/proposal/promulgation cycle; and (3) ensure that all technology contractors use a consistent approach.
 - Further supported by additional contractors with specific data collection and analysis assignments.

ESED is also responsible for:

- Source testing, in conjunction with IERL-RTP.
- Coordinating dispersion calculations with the Source-Receptor Analysis Branch, Monitoring and Data Analysis Division, OAQPS.

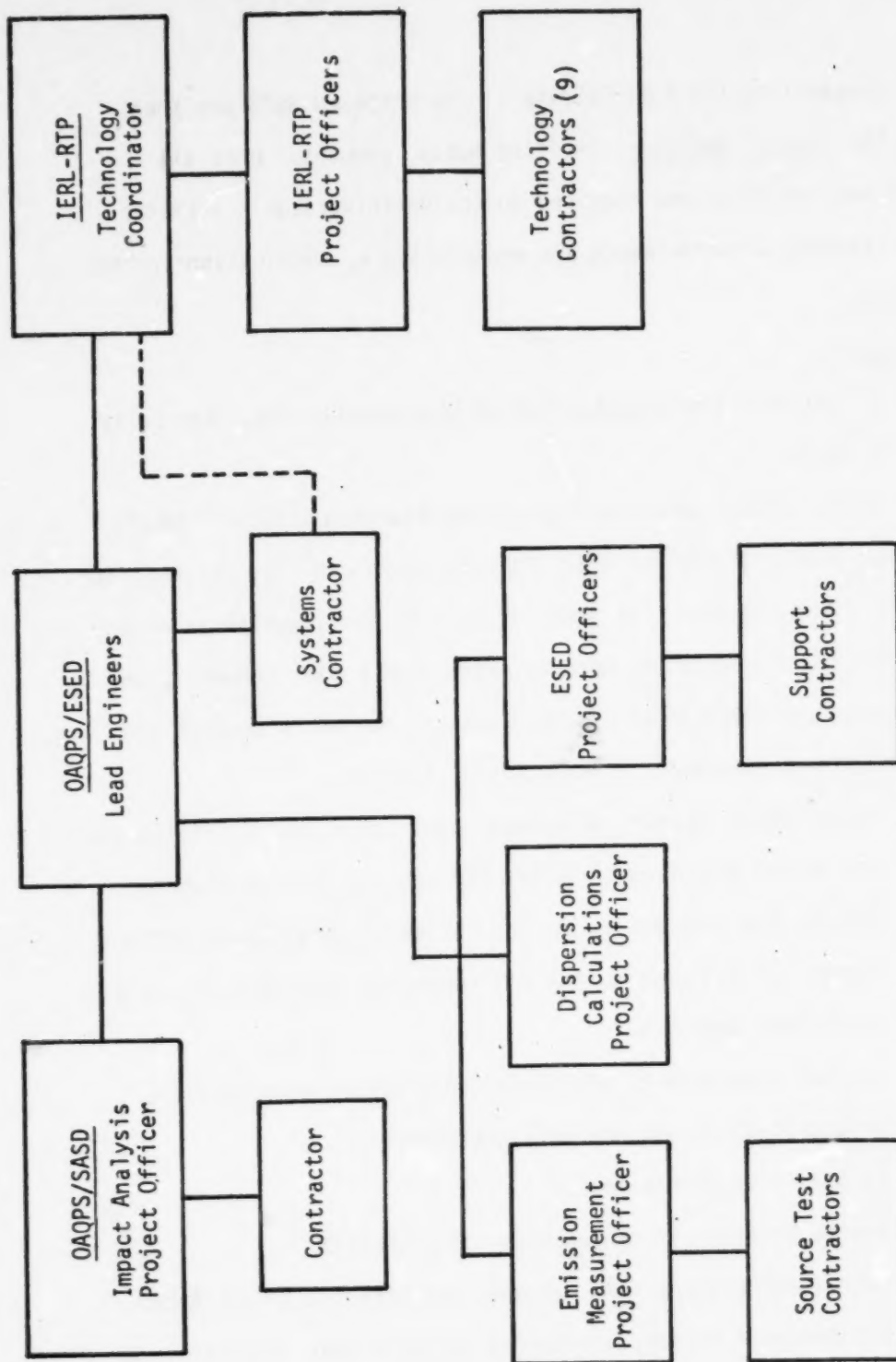


Figure 1. Project organization.

- IERL-RTP actively supports OAQPS on this project. In this role IERL-RTP:
 - Provides technical input on boiler characteristics and emissions and on available control technologies, including data on performance, costs, energy requirements, and other environmental impacts of the controls.
 - Assists in selecting alternative control options and model boilers.
- The Strategies and Air Standards Division (SASD) of OAQPS also supports the project actively by:
 - Providing analyses of the nationwide economic, environmental, and energy impacts associated with each alternative control.
 - Considering relationships between possible standards and fuel usage patterns.

Support by IERL-RTP was planned and arranged from the outset of the industrial boiler project. This approach utilizes the Laboratory's available expertise from ongoing R&D projects. Through its in-place contractual mechanism for obtaining outside support from contractors with control technology expertise, IERL-RTP is able to provide information to OAQPS in a timely fashion. This capability enables EPA to meet its stringent schedule for setting NSPS.

The effort to collect process and control technology information and prepare the BID is divided into the 15 tasks listed in Table 1. Figure 2 illustrates the interrelationships between these tasks by showing both the

Table 1. INDUSTRIAL BOILER TASKS

Task	Responsibility	Due Date
1. Boiler Classification	IERL-RTP	6/78
2. Emission Data	IERL-RTP/OAQPS	6/78
3. SIP's/Emission Estimates	OAQPS	9/78
4. Modifications	OAQPS	9/78
5. Fuel Switching	OAQPS	9/78
6. Cogeneration	OAQPS	9/78
7. Waste Firing	OAQPS	9/78
8. Model Boilers	IERL-RTP/OAQPS	2/79
9. Dispersion Analyses	OAQPS	6/79
10. Performance Tests	IERL-RTP/OAQPS	7/79
11. Control Technologies	IERL-RTP	8/79
12. Emerging Technologies	IERL-RTP	8/79
13. Fuel Use/Energy Impacts	OAQPS	8/79
14. Environmental Impacts	OAQPS	10/79
15. Economic Analysis	OAQPS	12/79

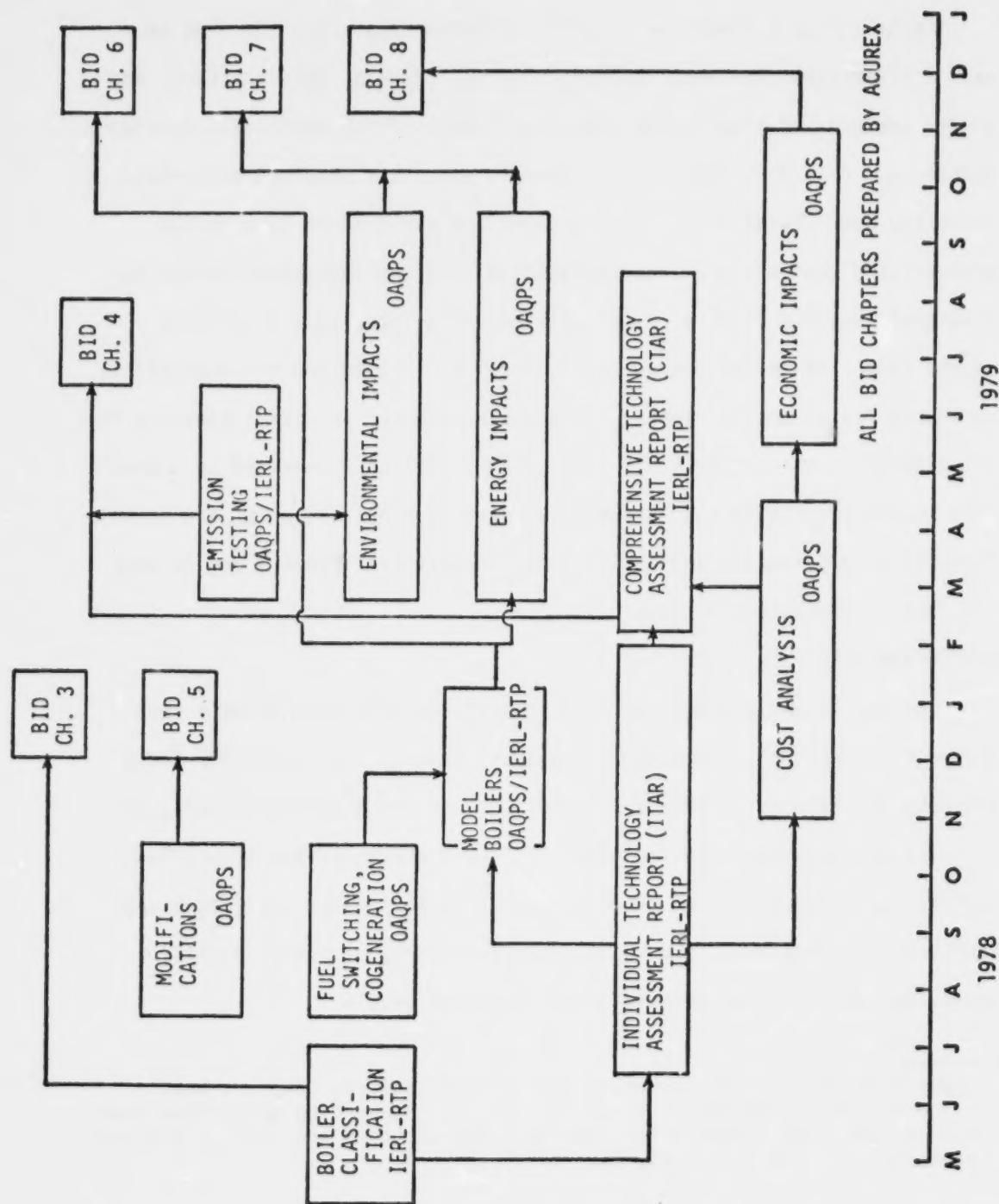


Figure 2. IERL-RTP/OAQPS integrated tasks.

timetable for completion of each task and the BID chapter which it supports. The key topics addressed in each chapter are listed in Table 2.*

Tasks 1-7 were completed in 1978 and define and scope the problem. Task 8 establishes the model boilers for the economic, environmental, and energy analysis of alternative standards. Most of the technical, environmental, and cost data that will be used to support a standard are being collected under Tasks 9-12. Descriptions and performance data on the demonstrated controls are being documented in eight Individual Technology Assessment Reports (ITAR's). As explained in a later section of this paper, each ITAR covers one control technology. Fluidized bed combustors, synthetic fuels, and/or cleaned fuels are considered potential controls for the purpose of this study. All eight ITAR's will be summarized by Acurex in a document entitled a Comprehensive Technology Assessment Report (CTAR). Tasks 13-15 provide for the analysis of the potential impacts due to each alternative regulatory option.

ISSUES ANALYSES

During the early stages of this project, several studies were conducted to scope the project and investigate issues. The tasks for these projects are listed in Table 1 as Tasks 1-7 and are described briefly here.

The boiler classification study of Task 1 characterized industrial boilers, provided installed capacity and projected growth data, developed baseline emissions data, and identified the effects of boiler design, operation, and fuel characteristics on emission levels.

* Figure 2 shows only the technology and impact analyses chapters that follow directly from the 15 tasks in Table 1. Supporting appendices have been omitted from Figure 2 for clarity. All of these are used to prepare Chapters 1, 2, and 9, thereby completing the BID.

Table 2. CHAPTERS IN A BACKGROUND INFORMATION DOCUMENT

1. Summary	Summarizes the recommended standard, with rationale, and estimates associated impacts.
2. Introduction	States the legislative basis and administrative process for this regulatory action.
3. Industrial Boilers	Provides a description of the source, its uses, current and projected population/emissions, and applicable SIP's.
4. Control Technologies	Describes the demonstrated and available controls and gives data on their emission reduction capability.
5. Modifications and Reconstructions	Describes possible changes to existing boilers to help decide, later, which should be considered modifications or reconstructions.
6. Best Systems of Emission Reductions	Develops model boilers with candidate best systems and alternative regulatory options to be analyzed in the following chapters.
7. Environmental Impacts	Identifies and discusses all environmental and energy impacts (beneficial and adverse) due to each alternative regulatory option.
8. Economic Impacts	Establishes and evaluates the expected economic impacts of each alternative regulatory option.
9. Rationale	Provides the rationale for the recommended standard.
Appendices	Contain test data and information on the history of the development of the standard, possible test methods, and enforcement issues.

Task 2 was a coordinated activity, involving both OAQPS and IERL-RTP, that determined the availability of sufficient data to support standards. As will be discussed later, EPA identified data gaps during this task and proposed a plan for conducting the necessary tests.

To evaluate the impact of emissions under each regulatory option relative to those that would occur if the boiler were required to comply only with a typical SIP level, Task 3 (SIP's/Emission Estimates) was included in this project. The object of this task was to define an average, or typical, SIP. This SIP level was also used to compute an average emission estimate for the U.S. population of boilers.

Since any existing boilers which are modified and reconstructed after the date of proposal of an NSPS must comply with the standard, Task 4 (Modification) was conducted to define the terms modifications and reconstructions for use in the standard writing procedure. In this task, the contractor identified likely changes to existing boilers that would be considered modifications or reconstructions according to the definition of those terms in the Code of Federal Regulations.

The effects on fuel usage patterns are an important consideration in this study. Therefore, Task 5 (Fuel-Switching) was issued to describe the technical factors involved in fuel switching and identify the likelihood of such changes among existing installations. To make this prediction, the task also estimated the cost and environmental impacts which result when industrial boilers currently burning gas or oil switch to oil, coal, or coal/oil mixtures.

Task 6 (Cogeneration) reports information on cogeneration, including an explanation of the concept of cogeneration, identification of those

industries most likely to employ cogeneration, emission levels from boilers used in cogeneration systems, applicable emission control techniques, and projected growth of capacity.

Task 7 (Waste-Firing) made a determination on the status of boilers firing industrial wastes. The result of this study was to establish waste-fired boilers as a separate NSPS activity.

CONTROL TECHNOLOGY ASSESSMENT

A key element in the process of developing an NSPS is the assessment of relevant demonstrated control technologies. Out of this assessment comes a list of candidate best systems, based on performance, commercial availability, adaptability to the process, etc. These candidate systems are then compared, and one or several may be selected as the basis for the standard, taking into account emission reduction capability, cost, energy penalty, and environmental side effects.

Standard Boilers

In this program the detailed assessment began with a survey of the industrial boiler population to better define and scope the study (see Task 1, above). Analysis of the distribution of installed boiler capacity and projected sales according to the boiler type, size, and fuel led to a classification of the members of this diverse category. A "standard" boiler was then selected to represent each class. Use of a classification scheme and "standard" boilers provided a common basis for all control technology contractors when assessing the overall performance of their assigned technology.

The following criteria were used to select the representative boilers:

- Extent of usage.
- Potential for uncontrolled emissions of particulate matter, SO_x , and NO_x .
- Representation of a cross section of the population.
- Projected sales.

Seven "standard" boilers (Table 3) were selected as representative of the industrial boiler population for use in the model boiler analysis. In addition, three coals were designated for the coal-fired boilers to represent the range of control requirements and associated impacts that could be expected from standards of performance. These are:

<u>Coal</u>	<u>Sulfur (%)</u>	<u>Ash (%)</u>	<u>Higher Heating Value (kJ/kg [Btu/lb])</u>
High sulfur eastern	3.5	10.6	27,450 (11,800)
Low sulfur eastern	0.9	6.9	32,100 (13,800)
Low sulfur western	0.6	5.4	22,300 (9,600)

Additional boilers and coals are being considered for use in an associated, more detailed computerized impact analysis.

Design parameters were also identified for each boiler to further ensure a common basis for reporting control technology capabilities. These parameters are: fuel rate; flue gas temperature and composition; load factor; excess air; and flue gas rate.

Individual Control Technology Assessments

For the technical and cost evaluations of the control techniques, IERL-RTP, with OAQPS agreement, divided the spectrum of possible technologies into eight generic types and assigned each one to a contractor with specific expertise in that technology. Table 4 lists these eight technolo-

Table 3. STANDARD BOILERS

Type	Fuel	Capacity MW (106 Btu/hr)
Package, firetube	Oil (distillate)	4.4 (15)
Package, firetube	Natural gas	4.4 (15)
Package, underfeed stoker, watertube	Coal	8.8 (30)
Field erected, chain grate stoker, watertube	Coal	22 (75)
Package, watertube	Oil (residual)	44 (150)
Field erected, spreader stoker, watertube	Coal	44 (150)
Field erected, pulverized coal, watertube	Coal	59 (200)

Table 4. TECHNOLOGY ASSESSMENTS

Project Title
1. Characterization of the Industrial Boiler Population
2. Clean Oil and Oil Treatment Technology
3. Coal Cleaning and Existing Clean Coal
4. Synthetic Fuels
5. Combustion Modification for NO _x Control
6. Fluidized Bed Combustion
7. Particulate Collection Technology
8. Flue Gas Desulfurization Technology
9. NO _x Flue Gas Treatment Technology
10. Comprehensive Technology Assessment Report (CTAR) and General Support

gies. It also shows the boiler population characterization study mentioned earlier and the task given to Acurex to help integrate these eight studies and compile their results into a single comprehensive report.

As noted earlier, the product of each control technology study will be an Individual Technology Assessment Report (ITAR). Within each ITAR the control systems that belong to that generic type are discussed separately. For example, the FGD report contains subsections dealing with lime/lime-stone, sodium scrubbing, Wellman-Lord systems, etc. To provide a complete evaluation of the capabilities of each control technology, the topics listed in Table 5 were addressed by every contractor.

Since control technologies operate over a wide range of emission reductions, three levels of control were identified for each technology and boiler/ fuel combinations at each level of control. In effect, the report for each technology documented the performance and impacts associated with the control of a single pollutant.

Information on each control technology is presented in the ITAR's according to the outline shown in Table 6. These ITAR's are currently being completed and will then be integrated into a Comprehensive Technology Assessment Report (CTAR) by Acurex. This overview document is intended to provide, in a single report, EPA's assessment of the state of the art in the control of NO_x , SO_x , and particulate emissions from fossil-fuel-fired industrial boilers. It will contain detailed discussions of individual technologies and of the systems that could be assembled to control all three pollutants simultaneously. The CTAR will discuss technologies that are in various stages of development and commercialization as well as those which have been demonstrated and are widely available.

Table 5. INDIVIDUAL TECHNOLOGY ASSESSMENT REPORT ISSUES

Status of development.

Applicability of control system to different boiler sizes and types.

Estimated capital and operating cost of the control system as a function of boiler size and type.

Control system cost as a function of removal capability.

Energy requirements of the control system.

Reliability of control system.

Environmental impact of waste streams.

Vendor availability.

Compatibility with other control systems.

Performance and operating data.

Table 6. OUTLINE OF THE INDIVIDUAL TECHNOLOGY ASSESSMENT REPORTS

-
-
- I. Executive Summary
 - II. Emission Control Techniques
 - III. Candidates for Best System of Emission Reduction
 - IV. Cost Analysis of Candidates for Best System of Emission Reduction
 - V. Energy Impact of Candidates for Best System of Emission Reduction
 - VI. Environmental Impact of Candidates for Best System of Emission Reduction
 - VII. Emission Source Test Data
-

The ITAR's have been used as the basis for selecting model boilers for the BID. These model boilers are a combination of "standard" boilers and candidate best control systems jointly accepted by OAQPS and IERL-RTP. The

"model boilers" are those which can be used by a boiler owner or operator to meet the prescribed emission level (i.e., one of the three discussed earlier), given a specified fuel.

The ITAR's and the CTAR form a partial basis for the control technology chapters of the Background Information Document (BID) prepared under the direction of OAQPS to support the proposed standard. The ITAR's and the CTAR contain much of the technology information that goes into the BID and provide in-depth coverage of all available and emerging control technologies. The BID, then, summarizes the key performance and operational characteristics of the demonstrated systems, stressing those which can be used by boiler owners under alternative regulatory options.

DATA GATHERING

Where data gaps exist, further performance testing is required. Testing is a joint activity of OAQPS and IERL-RTP. Testing requirements for data gathering were determined by summarizing the available test data and then assessing their quality. Criteria of acceptability included the presence of a complete process characterization of the boiler during the test, proper and representative operation of the boiler during the test, the use of Federal Reference Methods (FRM) to measure emissions, and a test of sufficient duration to assess the impact of coal variability. A matrix of available good quality data was established and compared to the data needed to support any probable standard at all levels, from moderate to stringent. That is, comparisons were used to determine, for example, if sufficient quality data existed to propose and defend an intermediate SO₂ standard for a medium sized stoker-fired boiler.

Priorities were set for filling in the data gaps on the basis of the expected need for that data, the extent of the data inadequacy, the cost of obtaining the data, and the availability of test sites. Tests were initiated in October 1978. Key tests in the series are: (1) continuous monitoring to obtain 30 days of data at boilers with flue gas desulfurization (FGD) and NO_x combustion modifications; (2) tests using FRM 5 at sites for particulate control of small oil- and coal-fired boilers; and (3) additional data on low Btu gasification, and fluidized bed combustion.

IMPACT ANALYSIS

Impacts which can be expected to result from the implementation of controls to meet alternative regulatory options are studied and discussed in the BID. These include potential environmental, energy, and economic consequences from these alternative actions. In the environmental analysis both primary impacts (e.g., those directly attributable to each control system, such as reduced levels of specific pollutants) and secondary impacts (e.g., indirect or induced impacts, such as exacerbation of another pollutant problem through utilization of a control system) will be identified and discussed. Primary emphasis will be placed on changes to the ambient air surrounding a typical source (NO_x , SO_x , and PM) and to changes in nationwide emissions, but attention will also be paid to other media (e.g., FGD sludge). Both beneficial and adverse effects will be assessed. The major emphasis will be on providing the reader with an accurate assessment of the incremental impact of the regulatory alternatives compared to sources which are uncontrolled or controlled to meet state regulations.

In addition, irreversible and irretrievable commitment of resources will be considered. As appropriate, this section will include a discussion

of the extent to which the alternative emission control systems may involve a trade-off between short-term environmental gains at the expense of long-term environmental losses, or vice versa, and the extent to which the alternative systems may result in irreversible and irretrievable commitment of resources.

The potential impact of an NSPS on energy consumption will be assessed by determining the amount and type (electric, fossil fuel, etc.) of energy required to operate each alternative emission control system. Where possible, these values will be compared to the quantities and types of energy required by a typical facility, both without emission control as well as controlled to comply with state regulations. Boiler-specific energy penalties will be extrapolated to estimate the impact of each regulatory option on increased national energy demands and on industry growth projections.

The economic impact of alternative control options will be evaluated during the preparation of the BID. To provide information for use in assessing the economic impact on a boiler owner of the cost associated with control options, a "business/economic" profile of the industry will be presented. Included in this discussion will be such factors as: industry structure, industry statistics, markets, prices, production and capacity, competition from imports, and other background information and data, as appropriate.

Following this industry overview, cost data will be presented for each emission control system. These costs will be given for new boilers and, separately, for existing units to cover modifications and reconstructions. Based on these results, the incremental costs associated with the alternative regulatory options will be identified. If the costs associated with

possible emission monitoring or compliance testing could be significant, these costs will be included in the overall costs. The costs will also consider those incurred to dispose of, in an environmentally acceptable manner, any liquid and solid wastes generated by air pollution controls.

The cost data described above will be used to evaluate the economic impacts associated with the incremental costs imposed on the source(s) to meet alternative regulatory options. The prime objective of the analysis and discussion will be to identify the incremental economic impact associated with alternative regulatory options.

This analysis will include the potential socio-economic and inflationary impacts that might result from the application of the alternative regulatory options.

In a typical BID the analysis is based on model plants (i.e., model boilers for this source category). Due to the complexity of the industrial boiler source category, a major portion of the analysis in this case will be performed by using a computer model. Included in these analyses of potential nationwide impact of various alternative regulatory options are the effects on fuel usage patterns--for example, the extent to which a standard on coal boilers might induce users to buy gas- or oil-fired boilers instead of coal-fired units. This analysis uses the Industrial Fuel Choice Analysis Model (IFCAM), a model originally developed for the Federal Energy Administration (now part of the Department of Energy--DOE) and now being modified for this project. This combination of approaches (i.e., models and computers) yields local impacts due to controls on representative samples of boilers plus nationwide impacts due to the application of controls to all affected boilers.

BID PREPARATION

The descriptions, analyses, and results described up to this point will be documented by Acurex for OAQPS in a Background Information Document (BID). This report will present a description of industrial boilers and the control technologies which have demonstrated their ability to control NO_x , SO_x , and PM from such boilers. It will also document the analyses of the potential economic, energy, and environmental impacts of the various alternative regulatory options that can be developed for these boilers. These analyses will assist EPA in arriving at a recommended standard, which is summarized in the BID and supported by a rationale.

SCHEDULE

The industrial boiler study was initiated in April 1978. It is scheduled for completion in August 1981, with promulgation of the standard by publication in the Federal Register. To date, draft ITAR's have been prepared, preliminary model boilers have been identified for internal review, and emission testing has begun. Presently targeted dates for the key remaining milestones are:

- BID completion, March 1980.
- National Air Pollution Control Techniques Advisory Committee meeting to be held, June 1980.
- Proposed standard published in Federal Register, October 1980.
- Promulgation of standard, August 1981.

Not included explicitly in this list are the numerous informal interactions planned with members from industry, DOE, environmental groups, and other government agencies.

SUMMARY

This paper has described a multifaceted EPA study that is being conducted with the intent of developing NSPS for industrial boilers. Two major organizations within EPA -- OAQPS and IERL-RTP -- are involved, as well as other EPA offices and support contractors. Through a series of tasks, background information is being collected on industrial coal-, oil-, and gas-fired boilers and on the technologies that have demonstrated the ability to reduce NO_x , SO_x , and PM emissions from these sources. This information collection activity is nearing completion and will be followed by extensive analyses of the potential economic, energy, and environmental impacts of alternative regulatory options. The data and analyses will be documented in a Background Information Document and used to develop and support an NSPS for the industrial boiler source category.

FLUE GAS DESULFURIZATION APPLICATIONS
TO INDUSTRIAL BOILERS

James C. Dickerman
Radian Corporation
Durham, North Carolina

INTRODUCTION

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Specifically mentioned in the Act are fossil fired steam generators. Accordingly, EPA has undertaken a study of industrial boilers with intent to propose standards of performance based upon the results of this and other studies.

This paper presents a summary of a study conducted to evaluate the applicability of various flue gas desulfurization (FGD) technologies for treating SO₂ emissions produced from industrial boilers. Results of this evaluation will be used by the EPA in preparing a NSPS for industrial boilers. Factors that were considered in evaluating the applicability of FGD technologies to industrial boilers included development status, capital and operating costs, energy requirements, environmental impacts, and performance and operating data.

APPROACH

In order to satisfy the objective of this study, a multiphased project approach was used. First, a comprehensive list of FGD processes was reviewed. This list included processes in commercial use, processes under development, and processes for which development efforts have ceased. Process status reports were prepared for eleven of the processes: those which are currently commercially used or are undergoing major demonstration efforts. Status reports for each candidate process contained detailed process descriptions, and discussions of design considerations and process performance characteristics.

The second phase involved selecting from the list of eleven candidate processes those that appeared to be best suited to industrial boiler applications.

Finally, detailed process analyses were prepared for the selected processes. Material and energy balance calculations were performed for each process to assess energy and environmental impacts as a function of boiler size, fuel sulfur content, and percent SO₂ removal. Also, a series of capital and operating cost estimates were made to assess the cost impact of each selected process as a function of boiler size, fuel sulfur content, and percent SO₂ removal.

SELECTION OF CANDIDATE PROCESSES

There are currently some 100 FGD processes that are in various stages of development, including processes in early developmental stages and those for which development efforts have ceased. Of these processes, there are five that are in commercial use today in the United States. In addition, there are six that are currently at the demonstration stage. It is felt that these eleven processes will be used for the majority of near-term FGD applications for both utility and industrial boilers. Table 1 presents a summary of the development status and applicability of these eleven processes to industrial boilers.

In order to select the candidate processes that appeared to be best suited for industrial boiler applications, a set of evaluation or screening criteria were established to provide an objective and consistent means of comparing the processes and to ensure that the same factors were considered for each process. The screening criteria were then applied to each process and the results were compared and used to select the processes that appeared to be best suited for near-term industrial boiler applications. The criteria used for this screening are listed in Table 2.

As a result of this screening step, four FGD processes were selected as candidate systems for application to industrial boilers. These processes were:

- Lime/Limestone
- Double Alkali
- Sodium Scrubbing
- Wellman-Lord

Each of these processes is commercially available and has been demonstrated to be an effective emission control technique. The first three techniques are currently being used to control SO₂ emissions from industrial boilers in the U.S. The Wellman-Lord process has been used for several other types of applications in the U.S., and on industrial boilers in Japan. A summary of the major characteristics of these processes is presented in Table 3.

The criteria having the greatest influence on this selection was status of development. The processes selected are better developed and have more acceptable operating histories than the remaining processes. It was felt that in order to provide support for new standards, FGD systems would have to be judged on their proven performance and not on what might be possible. The Citrate/Phosphate, Bergbau-Forschung/Foster Wheeler, Atomics International, and Shell/UOP processes were not selected because of their relatively undeveloped status. The magnesium oxide process, while operational on a full-scale unit, was not selected for evaluation because it has not yet operated continuously for longer than eight days.

In addition to the systems selected for further study, the Chiyoda 121 and the spray drying system have the potential for widespread application to both industrial and utility boilers. However, the development status of these two processes is such that there is insufficient data to permit the detailed analysis that is required to support an environmental standard. Because of the current interest in spray drying as an SO₂ control alternative, though, an independent analysis of that technology is being conducted as a continuation of this study.

ENERGY IMPACTS OF CANDIDATE CONTROL PROCESSES

Process energy requirements were evaluated as a function of process size, fuel sulfur content, and level of SO₂ control. Results of these calculations, shown below, indicate that the process energy penalties range from about 2 to 3½ percent of the gross heat input to the boiler for the three throwaway processes and from about 3 to 8 percent for the Wellman-Lord process. The larger energy consumption for the Wellman-Lord process is due to the steam and methane requirements for the regeneration and SO₂ reduction portions of the process.

RANGE OF FGD PROCESS ENERGY REQUIREMENTS

SO ₂ control method	Energy requirement (Percent of net heat input to boiler)
Lime/Limestone	2.6 - 3.7
Double-Alkali	2.0 - 2.3
Sodium Scrubbing	2.0 - 2.5
Wellman-Lord	3.2 - 8.0

The variations in energy requirements for these processes are due to different levels of sulfur in the coal, different levels of SO₂ control, and to a smaller extent, plant size.

A summary of the relative percentage of the energy requirements of each process area as compared to the overall energy requirement for the throwaway FGD processes is presented in Table 4. This table shows that stack gas reheat is the largest energy consumer for each of the throwaway processes. Methods of reducing the amount of energy required for flue gas reheat are currently under investigation by EPA. The primary method under consideration is bypassing a portion of the flue gas around the scrubber to heat the exit gas. However, the applicability of this method for installations requiring a high SO₂ control level is questionable.

The major energy consuming areas of the Wellman-Lord process vary depending upon the sulfur content of the coal being burned as shown in Table 5. For the eastern 3.5 percent sulfur coal case, which processes almost five times as much SO₂ as the western 0.6 percent sulfur coal case, the regeneration and sulfur production areas are the major energy users. For the low sulfur western coal case, stack gas reheat is the major energy consumer. It is doubtful that the energy requirements of the regeneration processing area can be significantly reduced since double effect evaporators were assumed for these calculations. Double effect evaporators are some 45 percent more energy efficient than single effect evaporators. As mentioned previously, methods of reducing stack gas reheat energy requirements are currently under investigation by EPA.

ENVIRONMENTAL IMPACTS OF CANDIDATE CONTROL PROCESSES

The air, liquid, and solid waste impacts of the candidate processes were considered as functions of size, SO₂ removal level, and fuel sulfur content. With regard to air pollution, each of the candidate FGD processes has the capability of both particulate and SO₂ removal, but only SO₂ removal was considered. The impact of all the candidate systems as far as SO₂ emissions is the same since each of the processes can be designed to achieve the same degree of SO₂ control.

Liquid Waste Impacts

With regard to water pollution, only the sodium throwaway process should produce a significant environmental impact. The other three systems can produce wastewater streams, but good design and operating practices can minimize any impact from these streams.

The aqueous waste stream from a sodium throwaway system will contain about five percent dissolved solids. In these systems, the absorbed SO₂ reacts to form Na₂SO₃ and Na₂SO₄, which are removed from the system as dissolved solids in an aqueous waste. Consequently, the amount of aqueous emissions is directly related to both the SO₂ control level and the coal's sulfur concentration. Discharge rates and average stream compositions for the cases considered in this study are given in Table 6.

Several existing water treatment technologies are potentially applicable for treatment of sodium throwaway FGD aqueous wastes. The other FGD processes should require no water treatment if good design practices are used. Technologies that may be used to reduce the level of total dissolved solids (TDS) in the sodium scrubbing system waste stream are: reverse osmosis, vapor compression distillation, and multistage flash evaporation.

Although these technologies may be technologically applicable for treating sodium throwaway process wastes, their overall complexity and cost may prohibit them from being used solely to treat aqueous wastes from industrial boiler FGD systems. It is likely that due to the small size of the discharge streams (Table 6), the small industrial boiler operators will be able to discharge this stream

into existing treatment facilities. The existing treatment facilities, however, will probably consist of the processes discussed above.

The major waste stream from a lime/limestone or double-alkali process is the sludge, which can contain significant quantities of supernatant liquid. However, good design and operating practices for the limestone and double-alkali processes include dewatering the sludge and recycling the supernatant liquid. Consequently, there should be essentially no water emissions from these systems except for times of severe rainfall or process upsets.

The aqueous waste stream from the prescrubber of the Wellman-Lord process will be characterized by a low pH which results from the chlorides that are removed from the gas stream. However, except for the high chloride concentrations and low pH, the quality of the prescrubber discharge will be very similar to that of the boiler ash sluice water. Since this stream has been estimated to be approximately one percent of the ash sluicing requirements for a power plant, it can be used for ash sluicing where it will become diluted and neutralized with the other ash sluice water. Consequently, water emissions from the Wellman-Lord prescrubber stream should be limited to intermittent discharges from the ash pond.

Solid Waste Impact

Of the four candidate processes only the limestone and double-alkali processes, both of which produce a sludge waste stream, would result in a major solid waste impact. A solids purge stream of Na_2SO_4 is produced in the Wellman-Lord process, but the stream is relatively small and should not constitute a major solid waste impact, especially for the size applications under consideration in this study.

Both the limestone and double-alkali sludges are composed primarily of calcium sulfite and sulfate salts. Significant amounts of fly ash may be present also, depending on the method of particulate control in use. The sludges are relatively inert and with proper site selection and proper disposal procedures, can be disposed of in an environmentally acceptable manner. The disposal methods currently in use are lined and unlined ponding and landfilling of treated and untreated materials. Potential adverse impacts of sludge disposal lie in the areas of disposal acreage requirements, water contamination through leaching and percolation of soluble

components of the solid waste into the groundwater system, and land use impacts due to poor structural properties. Treatment techniques to minimize adverse impacts may involve dewatering, addition of alkaline ash, and/or application of commercial stabilization technology. These techniques are used to decrease the sludge volume, decrease its permeability, and improve its structural properties.

As with the sodium throwaway system, the SO_2 absorbed from the flue gas by a double-alkali or limestone system must leave the process in a waste stream, in these cases as a waste sludge. Consequently, the amount of sludge produced is a function of unit size, fuel sulfur content, and the SO_2 removal level. Table 7 presents the results of the limestone process material balance calculations and shows the variation in sludge production with coal sulfur content and SO_2 removal.

The volume of sludge produced is also important as the sludge volume will determine the size of the holding pond or landfill area. Figure 1 illustrates the results of the sludge volume calculations graphically and shows the variation in sludge production with coal sulfur content, boiler size, and level of removal. Sludge volumes are presented in units of cc/sec, and acre-feet/15 years. The last category, acre-feet/15 years gives an indication of the total volume of sludge to be handled over the life of the plant assuming a 15 year life and an onstream factor of 60 percent.

COST IMPACT OF CANDIDATE PROCESSES

Process costs were evaluated as a function of process size, fuel sulfur content, and SO_2 removal for the four candidate FGD processes. The general approach used in developing the process costs consisted of four main steps. First, a series of material and energy balance calculations were performed for each process to define process stream rates and energy requirements as functions of unit size, SO_2 removal, and the amount of sulfur in the coal. Second, each of the FGD processes was divided into a number of process areas or modules, and economic scaling factors were developed for each process area based upon the type of equipment used. Third, detailed engineering design studies were used to obtain costs for each of the process areas. All of the FGD process costs were based on detailed design and economic estimates prepared by TVA except for the wastewater processing area used by the sodium scrubbing process. These costs were obtained from a report assessing wastewater control

technologies. Finally, the costs were individually scaled for each process area using the stream rates calculated in the material balances and the economic scaling factors developed for each process area.

Direct capital investment costs for the FGD processes ranged from a low of \$277,000 for a sodium throwaway process (8.8 MW, 75 percent removal, 0.6 percent sulfur) to a high of \$2,628,000 for a Wellman-Lord process (58.6 MW, 90 percent removal, 3.5 percent sulfur coal). When indirect capital expenses are added, the total capital investment costs for these two cases become \$691,000 and \$4,321,000 respectively. From a capital cost standpoint, for all cases considered the sodium throwaway process appears to be the least costly process, and the Wellman-Lord the most costly process.

With regard to annualized costs, the relative ranking of the various processes change with variations in unit size, SO₂ removal, and coal sulfur content. For the small unit size treating flue gas from a low sulfur coal (8.8 MW, 0.6% S) the sodium throwaway process appears to have the lowest annualized costs. However, for larger FGD system sizes treating flue gas from higher sulfur coals, the sodium process annualized costs increase faster than do the costs of the other processes. This is due to the relatively high cost of the sodium sorbent and costs associated with wastewater treatment. In fact, for the largest size considered (58.6 MW, 3.5% S coal, 90% removal) the sodium process had the highest annualized costs although it had the lowest capital investment costs. Figures 2 and 3 illustrate the relative capital and annualized costs of these processes graphically.

The cost effectiveness of the various FGD processes was also determined as part of this study. Cost effectiveness was defined as dollars per kilogram of removed SO₂ (\$/kg SO₂) and was calculated by dividing the annualized process costs by the kilograms of SO₂ removed in a year assuming a 60 percent load factor. Results of these calculations show that both coal sulfur content and process size significantly affect the cost effectiveness of an FGD process. For a given size system, cost effectiveness increases with an increasing coal sulfur content. For a fixed coal sulfur content, cost effectiveness increases with increasing process size. Consequently, the most cost effective systems are those designed for the 58.6 MW (200x10⁶ Btu/hr) boiler burning 3.5 percent sulfur coal, and the least cost effective systems are those designed for the 8.8 MW (30x10⁶ Btu/hr) boiler burning 0.6

percent sulfur coal. Figure 4 illustrates these effects for the lime/limestone processes. Curves developed for the other process showed similar effects.

Major findings of the cost analysis were:

- 1) An FGD system will add 10-50 percent to the uncontrolled boiler Total Capital Investment, depending upon the FGD system, boiler size, fuel sulfur content, and SO₂ control level. FGD annualized costs, including capital charges (15 yrs FGD life), will add 15-50 percent to uncontrolled boiler costs depending upon the same variables listed above.
- 2) It is less cost effective to remove SO₂ from flue gas with low SO₂ concentrations than from flue gas with high concentrations. Using an 8.8 MW sodium throwaway process as an example, it costs about 3½ times as much per unit of SO₂ removed (\$4.58/kg vs. \$1.33/kg) to remove 90 percent of the SO₂ from a 0.6 percent sulfur-based flue gas as from a 3.5 percent sulfur based gas.
- 3) The sodium throwaway process is estimated to have the lowest capital costs at all levels of operation. For small boiler sizes, and for low sulfur coal operations, the annualized costs of the sodium throwaway process are also lowest. However, for large high sulfur applications, the sodium throwaway process becomes the most costly alternative. This cost swing is due to costs associated with the sodium sorbent and the wastewater treating facility used for this study. For the large, high sulfur applications, either the lime/limestone or double-alkali process appear most attractive.

TABLE 1. FGD SYSTEM SUMMARY

Process	Development status	No. of operational plants		Applicability to industrial boilers
		Industrial	Utility	
Lime/Limestone	Commercial industrial and utility applications.	2	28	Generally applicable. Possible limitations due to solids disposal land requirements.
Double Alkali	Commercial industrial applications - Three utility applications are planned.	6	-	Generally applicable. Possible limitations due to solids disposal land requirements.
Wellman-Lord	Commercial applications for tail gas treating. A 115 MW utility demonstration test has been completed.	-	2	Generally applicable. Process costs and complexity may limit applications to only large boilers.
Magnesium Oxide	Commercial utility applications. No planned industrial applications.	-	1	Process complexity will limit applications for industrial boilers. Has not been operated continuously for longer than eight days.
Sodium Scrubbing	Commercial industrial and utility applications.	22	3	Generally applicable. Possible limitations due to sorbent availability and cost, and requirements for wastewater treatment.
Spray Drying	Pilot-scale. A 410 MW utility application is planned	-	-	SO ₂ removal may be limited for lime based high sulfur coal applications. System is generally applicable except for land requirements for solids disposal. High reliability is claimed but undemonstrated.
Citrate/Phosphate	1 MW pilot-scale. A 64 MW industrial boiler application is planned.	-	-	Applicability to small boilers will be limited by overall complexity and the need for a reducing gas to produce H ₂ S.

TABLE 1. (Continued)

Process	Development status	No. of operational plants		Applicability to industrial boilers
		Industrial	Utility	
Bergbau-Forschung/ Foster Wheeler	20 MW demonstration in U.S. and a 45 MW demonstration in Germany.	-	-	Applicability will be limited by overall complexity and the require- ment for extensive solids handling equipment.
Atomics International/ Aqueous Carbonate Process	1.25 MW nonintegrated pilot plant. A 100 MW utility demonstration is planned.	-	-	Applicability will be limited by overall complexity for small boiler applications. Use of unfamiliar technology in the reducing reactor may hinder process acceptability.
Shell/UOP	0.6 MW pilot plant in U.S. on coal-fired boiler. 40 MW in Japan on oil- fired boiler.	-	-	Applicability will be limited by overall complexity and the require- ment for hydrogen for regeneration.
Chiyoda 121	Small-scale pilot plant. A 20 MW utility demon- stration is planned.	-	-	Generally applicable. Possible limitations due to solids disposal land requirements in cases where by-product gypsum marketability is not feasible.

TABLE 2. FLUE GAS DESULFURIZATION SCREENING CRITERIA

1. Status of Development
 - Overall Process Development
 - Availability of Data
 2. Performance
 - SO₂ Removal
 - Reliability
 - Response to Flue Gas Composition Changes
 3. Applicability
 - Simplicity
 - Flexibility
 - Controllability
 - By-Product Marketability
 4. Economic Considerations
 - Capital Investment Costs
 - Operating Costs
 5. Energy Considerations
 - Liquid Pumping Requirements
 - System Pressure Drop
 - Regeneration Energy
 - Requirement for Reducing Gas
 6. Environmental Considerations
 - Multipollutant Control
 - Secondary Pollutant Emissions
-

TABLE 3. MAJOR CHARACTERISTICS OF CANDIDATE SYSTEMS

	SO ₂ Removal Efficiency (Percent)	Environmental Impacts	Relative Energy Impacts	Reliability	Development
Lime/Limestone	70 to >90	Calcium based sludge disposal	Medium	Utility systems had early problems, newer systems have shown improved reliability.	Commercial
Double-Alkali	>90	Calcium based sludge disposal	Low- Medium	High reliabilities reported, clear liquor system.	Commercial
Sodium Scrubbing	70 to >90	High TDS waste- water	Low- Medium	High reliabilities reported, clear liquor system with few process steps.	Commercial
Wellman-Lord	>90	Na ₂ SO ₄ solids	Medium- High	High reliability during utility demon- stration test. Clear liquor system although it has many process steps.	Commercial

TABLE 4. PERCENTAGE ENERGY CONSUMPTIONS FOR NONREGENERABLE PROCESSES
(58.6 MW Boiler, 90% Removal, Eastern Coal)

Source of energy consumption	Limestone		Double Alkali		Sodium TA	
	kW	Percent of total	kW	Percent of total	kW	Percent of total
Raw materials handling and preparation	114.2	6	14.5	1	14.5	1
Pumps	317.4	17	46.3	4	40.7	3
Fans	471.7	26	222.0	18	222	17
Reheat steam	884	48	884	73	884	66
Disposal	38.2	2	28.7	2	160.4	12
Utilities and services	8.6	<1	9.3	<1	9.3	<1
Total	1834.1		1204.8		1330.9	

TABLE 5. PERCENTAGE ENERGY CONSUMPTION FOR WELLMAN-LORD PROCESS
(58.6 MW, 90% SO₂ Removal)

Source of energy consumption	Western coal (0.6% S)		Eastern coal (3.5% S)	
	kW	Percent of total	kW	Percent of total
Raw materials handling and preparation	1.9	<1	9.3	<1
Pumps	42.2	2	82.4	2
Fans	208.6	11	205.0	5
Reheat steam	900	49	884	20
Process steam	469	25	2221	50
Methane	219	12	1048	23
Utilities and services	10.2	<1	9.9	<1
Total	1851		4460	

TABLE 6. WATER POLLUTION IMPACTS FOR THE SODIUM THROWAWAY SYSTEM*

Boiler size and type	Control level	3.5% S eastern coal		0.6% S western coal	
		lb/m ³	(gpm)	lb/m ³	(gpm)
8.8 kW (30 10 ⁶ Btu/hr) Underfeed Stoker	90	50.7	13.4	10.2	2.7
	85	47.3	12.5	9.8	2.6
	75	42.4	11.2	9.1	2.4
	56	31.2	8.4	-	-
22 MW (75 10 ⁶ Btu/hr) Chaingrate Stoker	90	130.9	34.6	27.6	7.3
	75	106.4	28.1	22.3	5.9
	56	75.3	19.9	-	-
44 MW (150 10 ⁶ Btu/hr) Spreader Stoker	90	262.7	69.4	55.3	14.6
	75	211.6	55.9	44.7	11.8
	56	146.8	38.8	-	-
58.6 MW (200 10 ⁶ Btu/hr) Pulverized coal	90	348.2	92.0	73.8	19.5
	85	315.6	83.4	66.6	17.6
	75	282.7	74.7	59.4	15.7
	56	200.2	52.9	-	-
Avg. Dissolved Solid Compositions		Na ₂ SO ₃	70 percent	Na ₂ SO ₃	52 percent
		Na ₂ SO ₄	18 percent	Na ₂ SO ₄	35 percent
		Na ₂ CO ₃	12 percent	Na ₂ CO ₃	13 percent
Avg. TDS Concentration (wt. %)		5.1		5.2	

*Based on material balance calculations provided in Appendix A.

TABLE 7. SOLID WASTE IMPACT FOR THE LIMESTONE FGD PROCESS
(Ash-Free Basis)

Boiler size and type	Percent removal	3.5% S eastern coal			0.6% western coal		
		g/s	lb/hr	l/min	gal/min	g/s	lb/hr
8.8 MW (30 10 ⁶ Btu/yr)	90	128.4	(1018)	5.0	(1.33)	29.9	(237)
Underfeed Stoker	85	121.2	(961)	4.7	(1.25)	27.7	(220)
	75	106.9	(848)	4.2	(1.10)	24.8	(197)
22 MW (75 10 ⁶ Btu/hr)	90	323.7	(2567)	12.9	(3.4)	74.6	(592)
Chafingrate Stoker	75	271.6	(2154)	10.6	(2.8)	62.2	(493)
44 MW (150 10 ⁶ Btu/hr)	90	652.9	(5177)	25.3	(6.7)	149.4	(1185)
Spreader Stoker	75	543.5	(4310)	21.2	(5.6)	124.6	(988)
58.6 MW (200 Btu/hr)	90	866.9	(6874)	33.7	(8.9)	199.3	(1580)
Pulverized Coal	85	823.0	(6526)	32.2	(8.5)	188.4	(1494)
	75	724.2	(5743)	28.4	(7.5)	165.8	(1315)

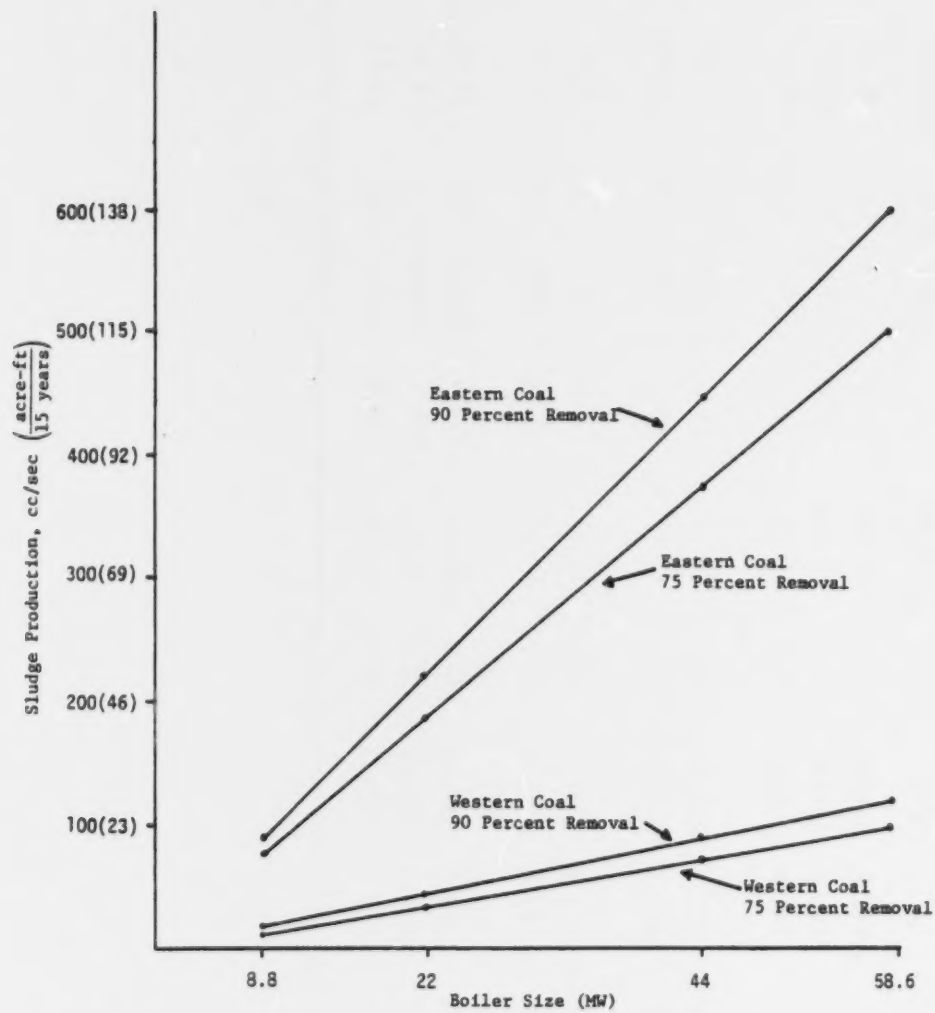


Figure 1. Sludge production rates for the limestone FGD process.

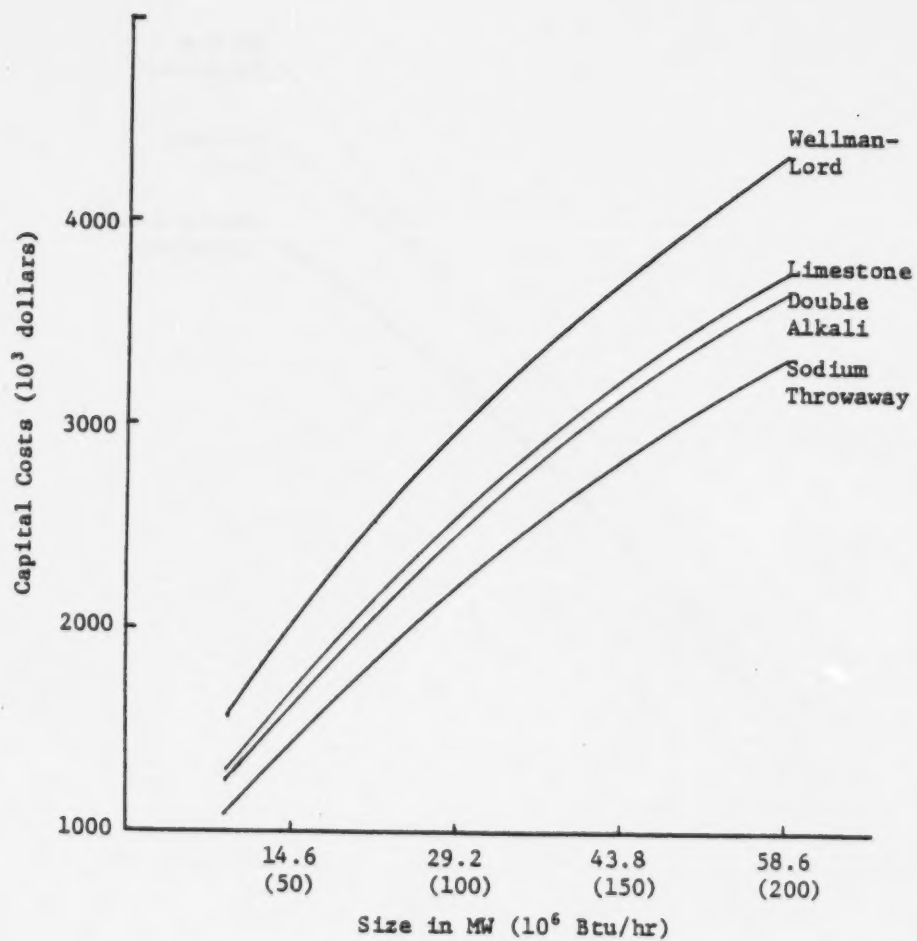


Figure 2. Capital costs versus unit size.
(3.5% S coal, 90% removal)

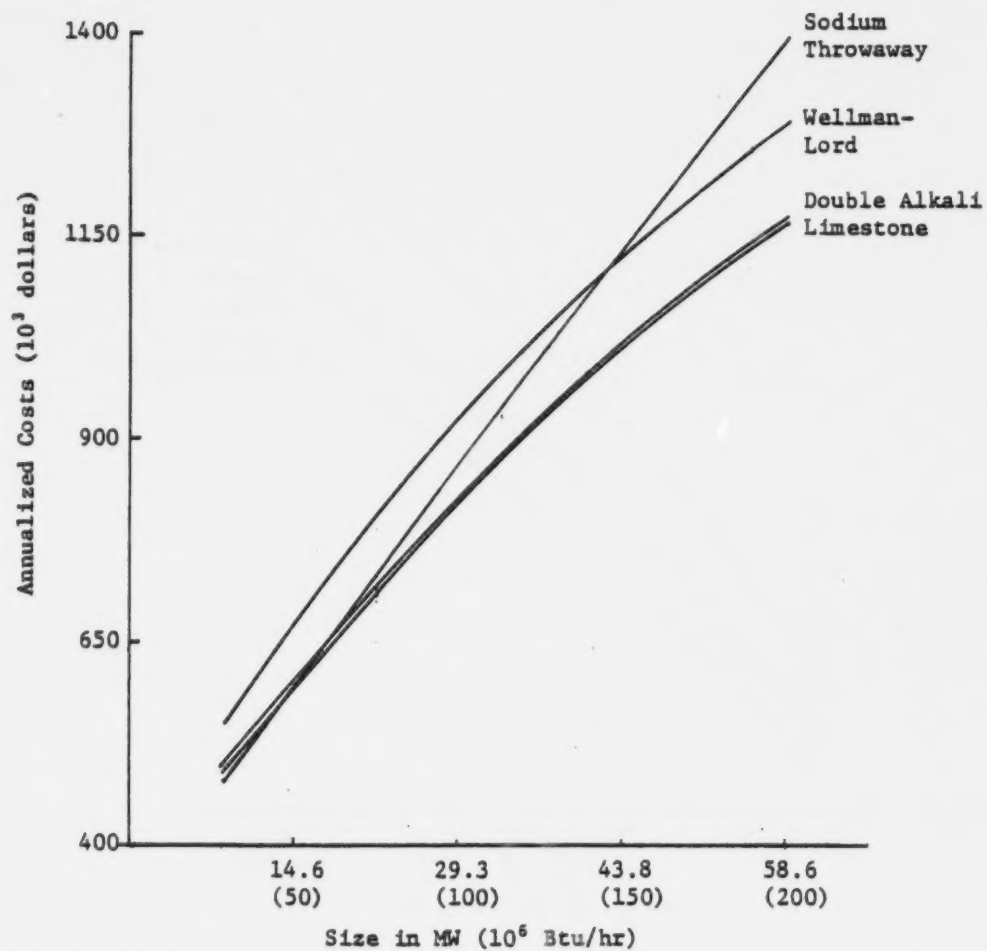


Figure 3. FGD annualized costs versus unit size.
(3.5% S coal, 90% removal)

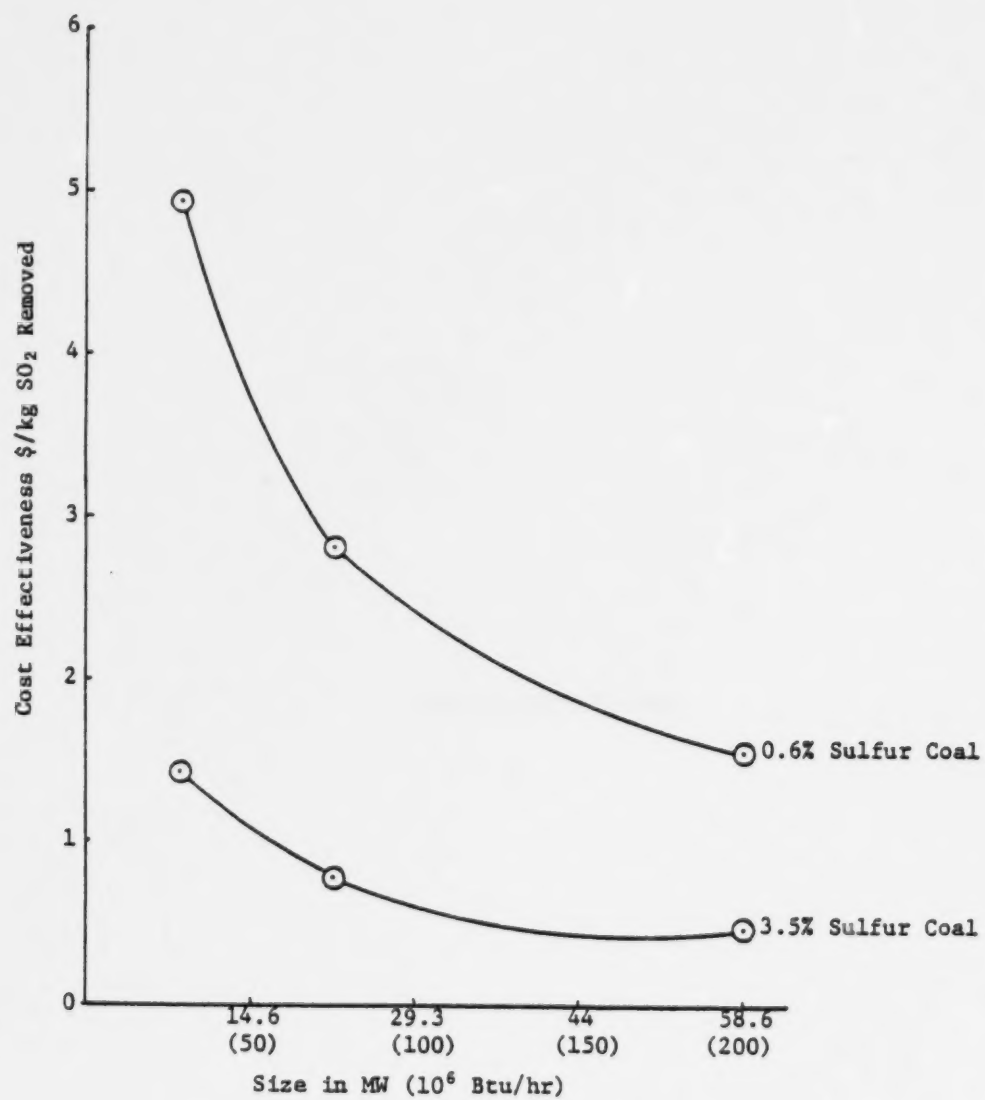


Figure 4. Limestone process cost effectiveness.

UNPRESENTED PAPERS

STACK GAS REHEAT - ENERGY AND ENVIRONMENTAL ASPECTS

Charles A. Muela
William R. Menzies
Radian Corporation
Austin, Texas

Theodore G. Brna
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina

ABSTRACT

Radian Corporation has completed a technical and economic assessment of stack gas reheat for the Environmental Protection Agency (EPA). The objectives of this study were to: 1) identify and analyze the potential problems associated with the generation of saturated flue gases by wet flue gas desulfurization processes, 2) determine current flue gas reheating practices in the utility industry, 3) determine how much reheat is actually required as well as its cost, and 4) compare the costs of the various techniques that can be used to reheat flue gases. Background data for this study were obtained from literature sources, from responses to questionnaires that were distributed to various users and suppliers of reheat equipment, and from visits to several facilities. In this paper, the potential benefits and corresponding energy requirements associated with the use of stack gas reheat are discussed.

Flue gas desulfurization (FGD) processes that cool and saturate flue gases with water may cause: 1) corrosion of downstream equipment, 2) a visible plume, 3) acid rainout in the vicinity of a plant's stack, and 4) increased ground-level concentrations of pollutants. Electric utilities, which are currently the primary users of FGD scrubbers, have cited equipment protection as the principal reason for the use of stack gas reheat. According to these users, the flue gas must normally be reheated by at least 16.7°C (30°F) to adequately protect equipment downstream of a wet FGD process.

It is extremely difficult to quantify the amount of reheat actually required in any given application because each of the four problems just cited is resolved by different levels of reheat. In general, however, the results of the analyses discussed in this paper indicate that reheating stack gases is an effective method of preventing the corrosion of downstream equipment, but it is less effective in significantly reducing the other potential problems.

INTRODUCTION

An EPA-sponsored study to assess the current state-of-the-art for stack gas reheat has been completed. Specific objectives of this study were to:

1. Identify and analyze the potential problems resulting from the generation of saturated flue gases by wet flue gas desulfurization processes.
2. Determine current practices in the utility industry regarding stack gas reheat.
3. Determine how much reheat is required to significantly reduce the potential problems caused by saturated flue gas.
4. Compare the costs of various reheat techniques.

The complete results of the analyses conducted to achieve these objectives are discussed in a draft report that is currently being reviewed by the EPA.¹ However, the scope of this paper is limited to an analysis of potential reheat-related problems and a determination of the energy requirements associated with significantly reducing those problems. Background information for the analyses presented in this paper and in the comprehensive study was obtained from the literature, questionnaires that were sent to companies in the utility industry and to vendors of reheat equipment, and visits to several power plants.

SO₂ emissions from utility plant boilers have been commonly controlled by flue gas desulfurization (FGD) processes. One type of FGD process, wet scrubbing, adds moisture to the flue gas and cools it to its adiabatic saturation temperature. A saturated flue gas may cause the following problems: 1) corrosion of equipment downstream of the FGD process, 2) a visible plume when the gas exits the stack, 3) acid rainout in the vicinity of the stack, and 4) increased ground-level concentrations of pollutants downwind from the stack.

The impacts of these problems can be reduced or eliminated by heating the flue gas above its saturation temperature. This can be achieved by several reheat configurations. The most commonly used are:

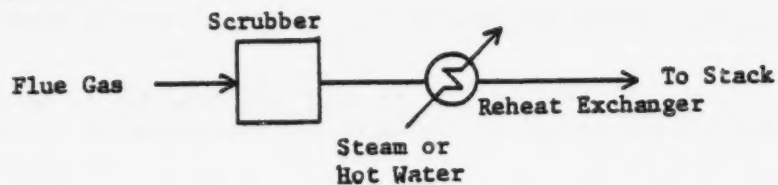
1. Inline Reheat (Figure 1a), in which the flue gas is heated by being passed through a heat exchanger downstream of the scrubber.
2. Indirect Hot Air Reheat (Figure 1b), in which heated air is mixed with the scrubbed flue gas.
3. Direct Combustion Reheat (Figure 1c), in which hot combustion gases generated by firing fuel oil or natural gas are mixed with the scrubbed flue gas.
4. Bypass Reheat (Figure 1d), in which a portion of the boiler flue gas is routed around the scrubber and mixed with the scrubbed flue gas.

The viability of bypass reheat is highly dependent on SO_2 standards. Because current and proposed SO_2 standards may restrict the use of bypass reheat in many applications, this reheat method is not considered in this paper. It should be noted, however, that bypass reheat has been used effectively at several facilities.

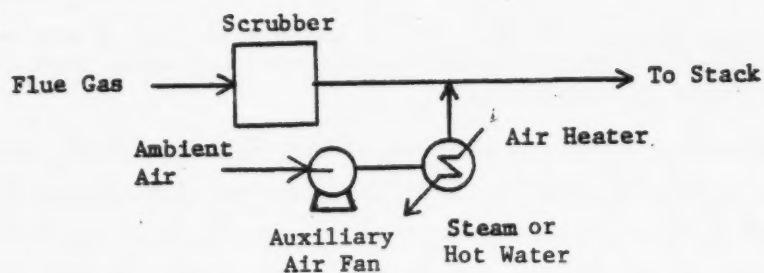
CURRENT PRACTICES IN UTILITY INDUSTRY

FGD process users and vendors, reheat users, and architect/engineer (A/E) firms were surveyed in order to identify current reheat practices in the utility industry as well as to determine the reliability of various reheat configurations. A profile of the different types of reheat configurations that have been or will be used in the utility industry was developed from data gathered in the survey and is presented in Table 1. These data indicate that inline reheat is the configuration utilities most prefer. The A/E's responding to the survey recommended indirect hot air be used when reheat is needed. However, most responding A/E's and vendors indicated that reheat is not necessary and, generally, do not recommend the use of reheat.

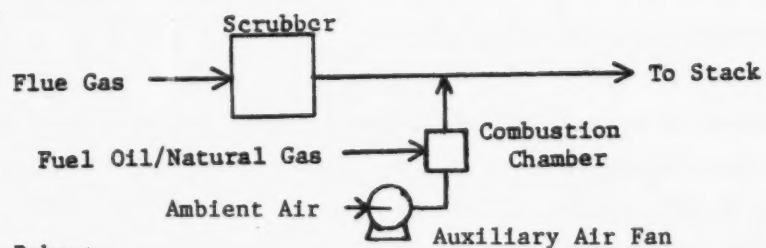
a) Inline Reheat



b) Indirect Hot Air Reheat



c) Direct Combustion Reheat



d) Bypass Reheat

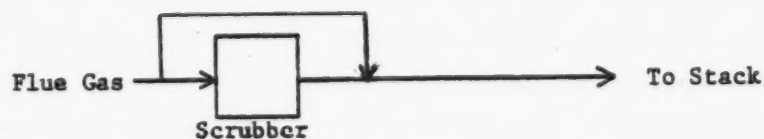


Figure 1. Simplified schematics of various reheat configurations.

TABLE 1. PROFILE OF VARIOUS REHEAT CONFIGURATIONS
USED WITH EXISTING OR PROPOSED UTILITY FGD SYSTEMS

Startup	Inline	Indirect Hot Air	Direct Combustion	Bypass	Wet Stack ^a
1968	1				
1969					
1970					
1971	2				
1972	4		2		1
1973	4		1		2
1974		3			
1975	1	1	1		1
1976	2	1	3		
1977	3	4	1	3	3
1978	3	2		7	4
1979	4	1	1	8	4
1980	3	2	3	7	3
1981	2				
1982	1				1
1983					1
Planned	<u>1</u>	<u>—</u>	<u>1</u>	<u>—</u>	<u>—</u>
TOTAL	31	14	13	25	20

^aWet stack--no reheat. Saturated flue gas is discharged directly to the atmosphere.

Although the need for reheat is site specific, utility companies responding to the survey indicated that reheat is generally used to prevent condensation and the corrosion of equipment. The degree of reheat used by these utilities varies from 0°C (0°F), corresponding to wet stack operation, to more than 55.6°C (100°F). A typical utility could be expected to reheat the flue gas by 22-33°C (40-60°F). The heating requirements corresponding to 27.8°C (50°F) of reheat in a 500-MW power plant were calculated for the inline, indirect hot air, and direct combustion reheat configurations. The results of these calculations are presented in Table 2. As the results in Table 2 indicate, the energy requirements associated with achieving this degree of reheat can be substantial, ranging from 1.6 to 2.8 percent of the total boiler heat input.

POTENTIAL BENEFITS AND ENERGY REQUIREMENTS ASSOCIATED WITH STACK GAS REHEAT APPLICATIONS

Each of the four problems that may be caused by saturated flue gases can be resolved with a different level of reheat. The impacts of stack gas reheat on these problems are examined below in terms of the:

- °benefits resulting from various levels of reheat, and
- °energy required to achieve these levels of reheat with several different reheat configurations.

The Use of Reheat for Equipment Protection

Scrubbed flue gases normally contain SO₂, SO₃, CO₂, chlorides, and sulfuric acid mist. Consequently, any water droplets in the flue gas will absorb these gases and become highly acidic. These acidic droplets can subsequently cause severe corrosion of equipment downstream of a wet FGD process. Moisture will be present when scrubbing liquor is entrained in the flue gas or when the flue gas temperature falls below its adiabatic saturation temperature and condensate forms. This temperature drop can

TABLE 2. HEAT INPUT REQUIRED BY VARIOUS REHEAT CONFIGURATIONS TO RAISE FLUE GAS TEMPERATURE 28°C (50°F)^d

	Inline	Direct Combustion	Indirect Hot Air
Temperature @ Scrubber Exit, °C (°F)	53.9 (129)	53.9 (129)	53.9 (129)
Temperature @ Stack Exit, °C (°F)	81.7 (179)	81.7 (179)	81.7 (179)
Reheat Air Characteristics			
Ambient Temperature, °C (°F)	-	-	15.6 (60)
Relative Humidity, %	-	-	50
Heated Air Temperature, °C (°F)	-	-	204 (400)
Flow Rate, 10 ⁶ kg/hr (10 ⁶ lb/hr)	-	-	0.60 (1.33)
Entrained Liquid ^a , 10 ⁻³ kg/std m ³ (gr/scf)	0.027 (0.012)	1.842 (0.805)	0.027 (0.012)
Heat Required for Vaporizing Entrainment MW (10 ⁶ Btu/hr)	0.038 (0.13)	2.64 (9.00)	0.038 (0.13)
Heat Required to Make Up Heat Losses ^b MW (10 ⁶ Btu/hr)	1.93 (6.60)	1.93 (6.60)	1.93 (6.60)
Sensible Heat Required for Reheat Medium ^c MW (10 ⁶ Btu/hr)	-	0.056 (0.19)	11.6 (39.6)
Sensible Heat Required to Raise Flue Gas Temperature 28°C (50°F), MW (10 ⁶ Btu/hr)	19.6 (66.8)	19.6 (66.8)	19.6 (66.8)
Total Heat Required, MW (10 ⁶ Btu/hr) (% of boiler input)	21.5 (73.5)	21.6 (73.7)	33.1 (113.1)
	1.63	1.64	2.51
	1.83	1.84	2.82

^aFrom the literature survey, the typical entrainment value for bottom wash mist eliminators was 0.012 gr/scf; for top wash mist eliminators, 0.805 gr/scf. (Ref. 2)

^bAssumed heat losses correspond to a 2.8°C (5°F) temperature drop which utilities have indicated is reasonable for duct work and stack.

^cThis sensible heat equals the heat required to raise the reheat medium from its ambient temperature to the stack exit temperature.

^dBases:

- 1) 500-MW Unit
- 2) 9503 kJ/kWh (9,000 Btu/kWh) Heat Rate
- 3) Forced Draft Fan Configuration
- 4) Flue Gas Saturation Temperature = 53.9°C (129°F)
- 5) Flue Gas Flow Rate = 2.33x10⁶ kg/hr (5.14x10⁶ lb/hr)
- 6) Flue Gas Heat Capacity = 1.09 kJ/kg-°C (0.26 Btu/lb-°F)
- 7) Direct combustion natural gas usage and an ambient air temperature of 15.6°C (60°F)

result from heat losses through the walls of the duct work and stack. A flue gas will typically cool about 2.8-5.6°C (5-10°F) in the duct work and stack following the scrubber.

Complete protection from corrosion of equipment downstream of a scrubber would involve the prevention of moisture and vaporization of sulfuric acid mist. The quantity of heat required to prevent the condensation is dependent on the total heat loss experienced by the flue gas, the amount of entrained scrubber liquor, and the type of reheat configuration used. The theoretical heat requirements of various reheat configurations to prevent the occurrence of moisture downstream of the scrubber were calculated and are compared in Table 3. In these calculations, only the forced draft fan arrangement of the various configurations was considered; consequently, no credit was taken for the work on compression by the fan (a fan downstream of a scrubber will raise the temperature of the flue gas). The data in Table 3 show that, although the indirect hot air configuration reduces the flue gas dew point, it requires about 23 percent more heat than inline reheat to eliminate moisture downstream of the scrubber. This increased heat requirement is a result of the energy required to heat ambient air to the flue gas dew point. The data in Table 3 also show that entrainment has a significant impact on the reheat requirement. Revaporization of sulfuric acid mist requires a considerably greater heat input than is needed to prevent condensation because the flue gas must be heated to the sulfuric acid dew-point temperature. In most systems, this temperature will be in the range of 93-149°C (200-300°F).

A comparison of Tables 2 and 3 indicates that the heat input required to raise the flue gas temperature 27.8°C (50°F) is significantly greater than the heat input needed to prevent moisture downstream of the scrubber. Consequently, the level of reheat generally used by the utility industry seems to be more than adequate to effectively protect downstream equipment from corrosion caused by components other than sulfuric acid.

TABLE 3. HEAT INPUT REQUIRED TO PREVENT MOISTURE DOWNSTREAM FROM SCHUBERT

	Inline			Reheat Configurations		
	Direct Combustion			Indirect Hot Air		
Entrained Liquid, 10 ⁻³ kg/Std m ³ (gr/scf) kg/hr (lb/hr)	0.027 59.0	(0.012) 4010	1.842 (0.805) (8840)	0.027 59.0	(0.012) 4010	1.842 (0.805) (8840)
Dew Point Temperature at Stack Exit, °C (°F)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)
Ambient Air Characteristics						
Temperature, °C (°F)	-	-	-	-	-	-
Relative Humidity, %	-	-	-	-	-	-
Heated Air Temperature, °C (°F)	-	-	-	-	-	-
Flow Rate, 10 ⁶ kg/hr (10 ⁴ lb/hr)	-	-	-	-	-	-
Assumed Heat Loss ^a , MW (10 ⁶ Btu/hr)	1.93 (6.60)	1.93 (6.60)	1.93 (6.60)	1.93 (6.60)	1.93 (6.60)	1.93 (6.60)
Heat Required for Vapor- izing Entrainment MW (10 ⁶ Btu/hr)	0.036 (0.13)	0.036 (0.13)	0.036 (0.13)	0.036 (0.13)	0.036 (0.13)	0.036 (0.13)
Sensible Heat Required for Reheat Medium ^b MW (10 ⁶ Btu/hr)	-	-	-	0.003 (0.01)	0.006 (0.02)	0.47 (1.6)
Theoretical Heat Required to Prevent Moisture, MW (10 ⁶ Btu/hr)	2.0 (6.7)	2.0 (6.7)	2.0 (6.7)	2.0 (6.7)	2.0 (6.7)	2.0 (6.7)
(% of boiler input)	0.15	0.15	0.15	0.15	0.15	0.15

^aTotal heat losses are assumed to be equivalent for all configurations and correspond to a 2.8°C (5°F) drop in flue gas temperature.

^bThis sensible heat equals the heat required to raise the reheat medium from its ambient temperature to the stack exit temperature.

- ^cBases:
- 1) 500-MW Unit
 - 2) 9503 kJ/kWh (9,000 Btu/kWh) Heat Rate
 - 3) Forced Draft Fan Configuration
 - 4) Flue Gas Saturation Temperature = 53.9°C (129°F)
 - 5) Flue Gas Flow Rate = 2.33x10⁶ kg/hr (5.14x10⁶ lb/hr)
 - 6) Flue Gas Heat Capacity = 1.09 kJ/kg-°C (0.26 Btu/lb-°F)
 - 7) For direct combustion, natural gas ambient temperature is assumed to be 15.6°C (60°F)

Suppression of Visible Plume

The use of a wet scrubber can result in a visible plume. While a visible plume does not have a direct negative impact on the environment, it can be aesthetically displeasing, potentially hazardous to ground and air traffic, and in violation of air pollution control ordinances restricting opacity. The mechanics of visible plume formation are illustrated in Figure 2. Ambient air conditions (temperature, relative humidity) are represented by point 1 in this figure. Point 2 corresponds to the conditions of a hot flue gas as it exits the boiler. The flue gas is saturated and cooled during the scrubbing operation and is represented by point 3. As a saturated flue gas leaves the stack, it mixes with ambient air according to line 3-1. A visible plume is formed when the ambient-air-saturated flue gas mixture intersects the saturation curve and crosses into the fogged field area of the chart.^{3,4} The persistence of the visible plume increases with increasing length of the portion of line 1-3 in the fogged field.

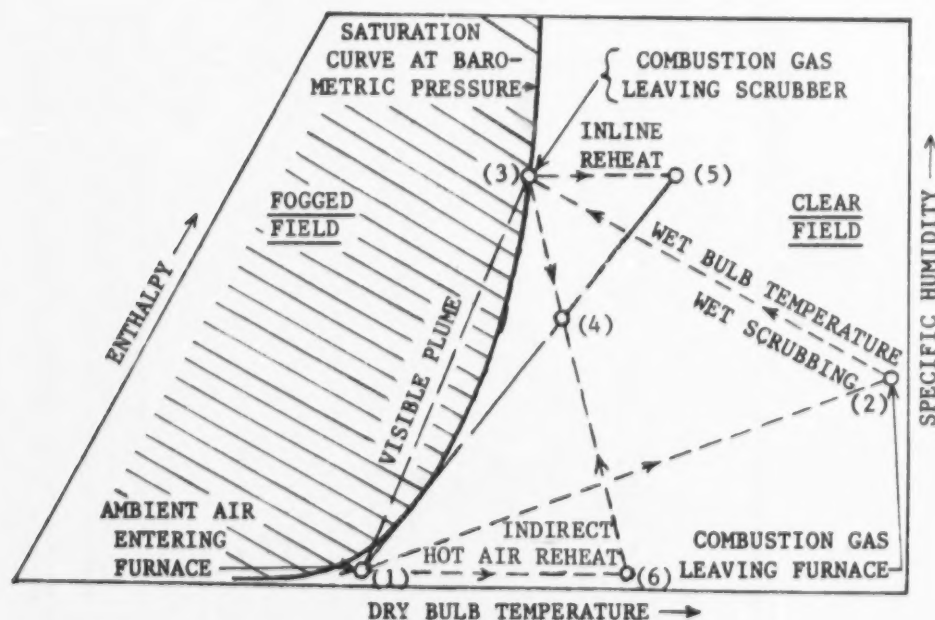


Figure 2. Psychrometric chart showing state points of flue-gas/air mixture with scrubbing and reheat. (Ref. 4,5)

Prevention of a visible plume with reheat involves the clockwise rotation of the saturated flue-gas/ambient-air mixing line (line 3-1, Figure 2) until it is tangent (line 1-5) to the saturation curve. The temperature to which the flue gas must be heated in order to prevent a visible plume is represented by point 5 when inline reheat is used and by point 4 when indirect hot air reheat is used. Note that the degree of reheat for inline reheating (the difference in dry bulb temperatures at points 5 and 3) is greater than for indirect hot air reheat (dry bulb temperature at point 4 less that at point 3) shown in Figure 2. Due to the shape of the saturation curve, the temperature of the heated flue gas and the corresponding reheat requirements are dependent on the ambient air's temperature and relative humidity. The reheat temperature (temperature of the scrubbed flue gas after heating) and the theoretical heat input requirements of the inline and indirect hot air configurations at various ambient air conditions are shown in Table 4. (The flue gas saturation temperature in Table 4 corresponds to point 3, while point 6 of Figure 2 represents the dry bulb temperature of hot air after indirect reheat.) The data in Table 4 show that the required reheat temperature (point 5 or 4) and heat input are highly dependent on the ambient air condition (point 1). These data also show that indirect hot air reheat, via dilution of the flue gas, significantly decreases the reheat temperature required by inline reheat. Figure 2 also illustrates this: the temperature at point 4 is less than at point 5. However, the addition of air increases the total mass of gas which must be heated. As a result, the theoretical heating requirements for the two reheat configurations are approximately equal.

When a forced draft, indirect hot air configuration is used (Figure 1b), the work of compression from the auxiliary air fan ultimately helps to raise the temperature of the flue gas; this reduces the quantity of heat that must be added to the flue gas to achieve the required reheat temperature. These reduced heat inputs are shown in Table 4. As the data in Table 4 indicate, the prevention of a visible plume could be very costly in terms of the required heat input.

TABLE 4. ENERGY REQUIREMENTS ASSOCIATED WITH PREVENTION OF A VISIBLE PLUME^a

Reheat Configuration Ambient Air Temperature, °C (°F) Ambient Air Relative Humidity, %	Inline			Indirect Hot Air		
	15.6 (60)	50	100	15.6 (60)	50	100
Flue Gas Saturation Temperature, °C (°F)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)	53.9 (129)
Heated Air Temperature, °C (°F)	-	-	-	204 (400)	204 (400)	204 (400)
Quantity of Heated Air Required, 10 ⁶ kg/hr (10 ⁶ lb/hr)	-	-	-	0.400 (0.881)	0.798 (1.76)	2.03 (4.47)
Flue Gas Reheat Temperature Required to Prevent Visible Plume, °C (°F)	83.9 (183)	115.6 (240)	226.1 (439)	74.4 (166)	91.1 (196)	122.8 (253)
Reheat Required to Prevent Visible Plume Formation, MW (10 ⁶ Btu/hr)	20.8 (71.0)	43.7 (149.0)	121.9 (416.0)	20.8 (71.0)	43.7 (149.0)	121.9 (416.0)
(% Boiler Input)	1.58	3.31	9.24	20.6 ^b (70.3)	43.2 (147.7)	121.0 (412.7)
				1.58	3.31	9.24
				1.56	3.28	9.17

^a Bases and Comments: 1) Flue gas is representative of a 500-MW plant.

2) Heat losses in duct work and stack are assumed to be negligible.

3) Entrainment is assumed to be zero.

4) Forced draft fan arrangement.

^b Underlined reheat requirements were developed for indirect hot air by taking credit for work of compression produced by the auxiliary fan (see Figure 1b). The pressure drop was assumed to be 6 in. H₂O and an 85 percent fan efficiency was also assumed.

If the flue gas saturation temperature in Table 4 is raised by 27.8°C (50°F), the resulting reheat and exit temperature would be 81.7°C (179°F). Comparing this temperature with those flue gas temperatures that are required to prevent a visible plume indicates that the amount of reheat typically used by industry will not prevent visible plumes at many meteorological conditions.

Using Reheat to Prevent Acid Rain

When a wet FGD process is used, SO_2 , SO_3 , and water in the flue gas can react to form H_2SO_3 and H_2SO_4 . Because the dew point of sulfuric acid is normally higher than the adiabatic saturation temperature of the flue gas, the H_2SO_4 vapor which is formed can condense even when the flue gas is above its saturation temperature. The resulting acid mist may, in turn, provide the nuclei for additional condensation, and rainout may occur when the flue gas leaves the stack. Acid rain may also be formed when residual SO_2 is oxidized and absorbed by moisture on the stack wall and the resulting H_2SO_3 and or H_2SO_4 droplets are entrained in the flue gas leaving the stack.

Reheat can suppress acid rain by:

- °preventing condensation of water vapor in the system,
- °vaporizing any entrained liquid leaving the mist eliminator, and/or
- °revaporizing sulfuric acid mist that is present in the system.

As mentioned previously, preventing the condensation of water vapor and eliminating any entrainment that may be present require the input of only enough heat to keep the scrubbed flue gas above its dew point (about 52-60°C (125-140°F) in most systems), while revaporizing any sulfuric acid that is present requires the flue gas to be heated to

above the sulfuric acid dew point (approximately 93-149°C (200-300°F) in most systems, although this temperature is dependent on H_2O and SO_3 concentrations in the flue gas). Note that, even though all the sulfuric acid in the system can be revaporized with a substantial heat input, the condensation and subsequent rainout of sulfuric and/or sulfurous acids could still occur when the scrubbed flue gas mixes with cooler ambient air.

The impact of reheat on the potential for rainout from a wet plume was simulated to determine the impact that reheat could have on the water concentration in a plume and the time period over which this concentration was between 0.5 and 1 g/m³. Blum⁶ suggested this as the critical concentration range that would result in rainout from a plume. This analysis indicated that reheat can have a significant impact on the potential for rainout from a plume at conditions of mild temperatures and low relative humidities. However, since the mechanism causing rainout is not fully understood, this conclusion must be considered to be somewhat speculative.

Impacts of Scrubbing and Reheating on Ground-Level Pollutant Concentration

Although wet FGD processes remove SO_2 from the flue gas, they will increase the ground-level concentrations of other pollutants (such as NO_x) by reducing plume buoyancy and rise. Reheating the flue gas increases buoyancy of the gas and thus reduces the ground-level pollutant concentrations exhibited by the scrubbed flue gas. The effect of scrubbing and reheating a flue gas on the short term (3-hour) SO_2 and NO_x ground-level concentrations was analyzed. Two atmospheric stabilities, unstable and neutral, and three levels of reheat were considered in the analysis. An unstable atmosphere, although occurring infrequently, is known to produce high ground-level pollutant concentration. A neutral atmosphere does not produce as high a ground-level concentration as the unstable atmosphere, but it occurs more frequently.

The results of the unstable atmosphere analysis are presented for SO_2 and NO_x in Figure 3, which shows that:

°The highest SO_2 and lowest NO_x ground-level concentrations are exhibited by the unscrubbed flue gas. These results reflect the ability of the scrubbing process to remove SO_2 but not NO_x .

°Scrubbing the flue gas reduces the maximum unscrubbed ground-level SO_2 concentration by approximately 48 percent. However, scrubbing increases the maximum unscrubbed ground-level NO_x concentration by about 160 percent. It should be noted that for both SO_2 and NO_x the predicted concentrations are below applicable ambient air quality standards.

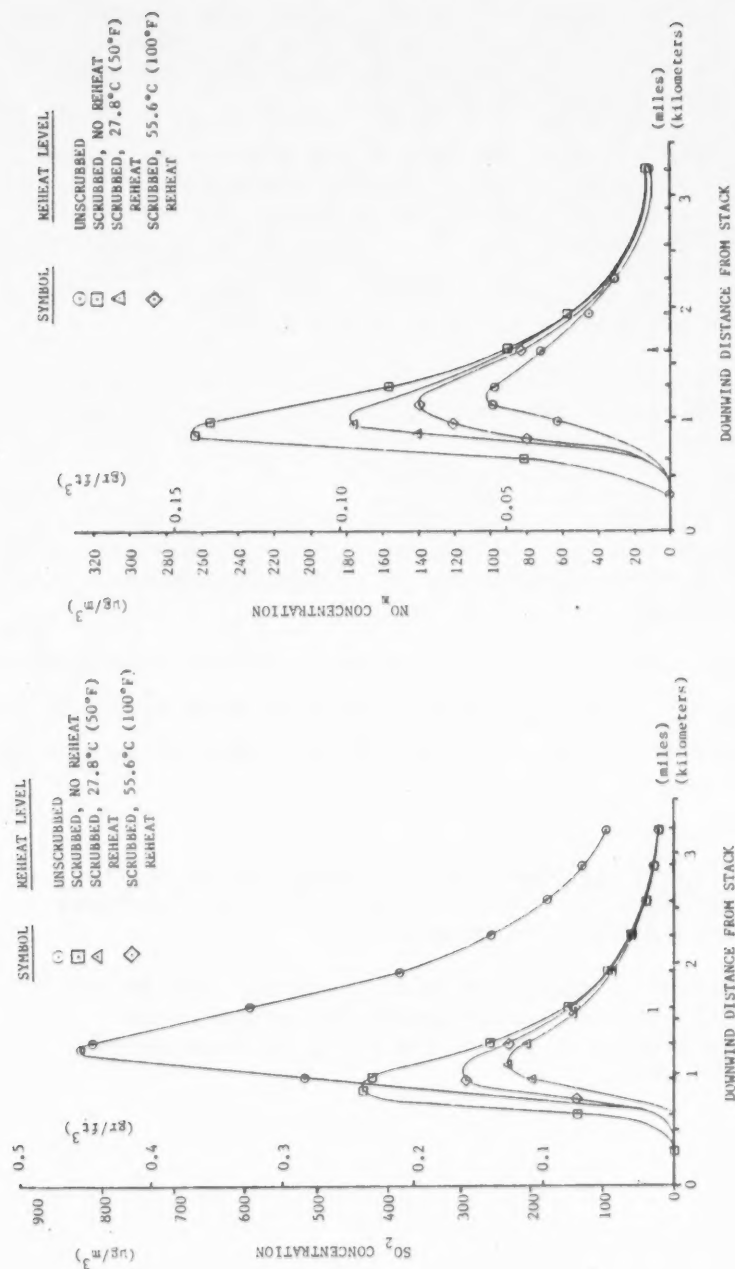
°Reheating the scrubbed flue gas 27.8°C (50°F) reduces the maximum ground-level SO_2 and NO_x concentrations by about 33 percent compared with scrubbing with no reheat.

°The addition of 55.6°C (100°F) of reheat to the scrubbed flue gas reduces the maximum ground-level SO_2 and NO_x concentrations by 47 percent over a scrubbed but unreheated flue gas.

The use of reheat under neutral atmospheric stability conditions produces similar trends in SO_2 and NO_x concentrations as those predicted for an unstable atmospheric condition. From this analysis, it is concluded that:

°Reheating a scrubbed flue gas significantly reduces the ground-level pollutant concentrations attributed to the unreheated scrubbed gas.

°A substantial degree of reheat (greater than 55.6°C (100°F)) is required to reduce the ground-level concentrations of NO_x to its original (unscrubbed) concentration.



Bases:

- 1) 32 sulfur coal
- 2) Ambient air temperature, 15.6°C (60°F)
- 3) 80% SO₂ removal
- 4) Vertical temperature gradient, -19.5°C/km (-10.7°F/1000 ft)
- 5) Flue gas exit temperature and velocity:

	Temperature °C (°F)	Exit Velocity m/s (ft/sec)
Unscrubbed	149 (300)	10.7 (35)
Scrubbed, No Reheat	53.9 (129)	8.75 (28.7)
Scrubbed, 27.8°C (50°F) Reheat	81.7 (179)	9.50 (31.1)
Scrubbed, 55.6°C (100°F) Reheat	109 (229)	10.2 (33.5)

Figure 3. Three-hour, ground-level SO₂ and NO_x concentrations downwind of stack (unstable atmosphere, wind speed = 2.24 m/s (5 mph)).

The impact of the inline and indirect reheat configurations on ground-level pollutant concentrations was compared. This comparison showed that for the same heat input the two reheat configurations produced similar maximum ground-level pollutant concentrations. When the scrubbed gas is heated to the same temperature with the two configurations, the indirect hot air configuration produces the lower maximum ground-level concentration.

SUMMARY

The major points of the information presented in this paper are:

°Currently, the utility industry utilizes stack gas reheat primarily to protect equipment from corrosion. The degree of reheat used varies from 0-55.6°C (0-100°F), with the average being about 27.8°C (50°F). The heat input required to reheat the scrubbed flue gas from a 500-MW power plant through 27.8°C (50°F) can vary from approximately 1.6 percent of the boiler input for the inline reheat to about 2.8 percent for the indirect hot air configuration.

°Moisture in the flue gas enhances the corrosion of equipment downstream of the scrubber. For the cases and reheat configurations analyzed in this paper, the heat input required to prevent moisture downstream of the scrubber varies from 0.2 to 0.4 percent of the boiler input. The inline reheat configuration requires the least heat input in preventing moisture downstream of the scrubber.

°The heat required to prevent a visible plume is about the same regardless of the reheat configuration used. Prevention of a visible plume is very dependent on meteorological conditions, and reheat cannot prevent a visible plume with a reasonable degree of reheat at many ambient air conditions.

°At present, knowledge about the mechanisms of acid rainout is limited. Consequently, the impact of reheat on acid rainout cannot be fully evaluated. However, reheat can certainly lessen the potential for acid rain by preventing moisture downstream of the scrubber.

°Stack gas reheat can significantly reduce the ground-level pollutant concentrations from a scrubbed flue gas. Although the lowest pollutant concentrations are produced by high levels of reheat, an incremental increase in the reheat level does not produce a proportional decrease in pollutant concentration.

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Minimizing Operating Costs of Lime/Limestone FGD Systems

Carlton Johnson
Manager of Process Engineering
Peabody Process Systems

Flue gas desulfurization systems all too frequently are put in the category of black magic. Statements are made that sophisticated controls are required and very specific conditions set to make SO₂ scrubbing systems perform as required. Nothing could be further from the truth. It is Peabody's experience gained in design, construction and operating results over a period of years that such exact considerations are not required and that the system, if properly designed originally, can operate under a wide variety of conditions. We must emphasize "if properly designed originally".

Selection of Design Criteria

When selecting the design criteria for its FGD System, a utility must allow for a wide variety of conditions. Consideration must be given to the maximum percent sulfur in the coal that is to be burned. Allowance must be made for the increased leakage of air into the system as the power plant ages. Contingency may also be allowed for the fact that it may be possible to debottleneck the generating system beyond the rated capacity. The combination of the above factors will result in design conditions for the FGD System in excess of what would be required to meet normal operating conditions. In fact, such design conditions may never be reached during the life of the generating unit, but are required to insure meeting emission standards under all conditions. The margin between the design and normal operating conditions will ensure that environmental regulations can be met; however if a higher operating cost is then necessary, while it is necessary to meet the environmental regulations, to exceed them provides no benefit to the utility.

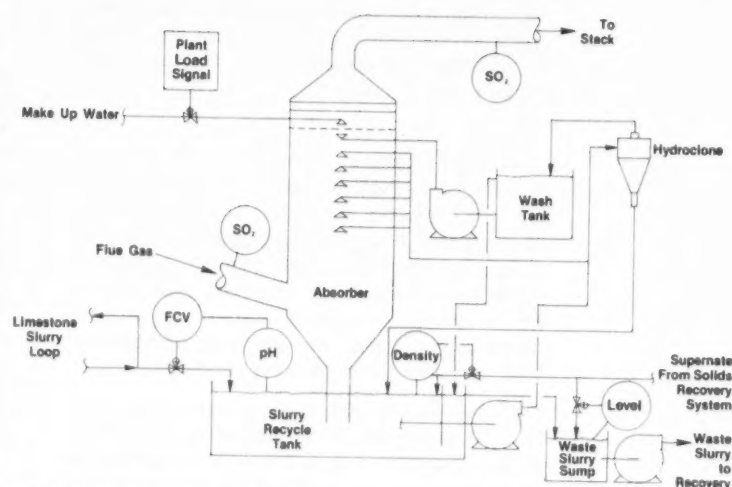


Figure 1

What Comprises Operating Costs?

When establishing design criteria, the owner and engineer must be aware of the major factors which contribute to operating costs of the FGD system. These are:

1. Alkali Consumption
 - a) Quantity of SO₂ removed
 - b) Stoichiometry
2. Sludge
 - a) Quantity of SO₂ removed
 - b) Alkali stoichiometry
3. Power
 - a) L/G ratio (GPM/1000 ACFM)
 - b) Gas System Pressure Drop

Alkali consumption and waste sludge production comprises the largest part of the operating cost. Each of these costs is dependent upon the amount of SO₂ removed and the stoichiometry or efficiency of alkali utilization at which the SO₂ is removed. A poor utilization of alkali means that not only is the alkali consumption higher but also that unreacted alkali is contained in the waste sludge and it is disposed of with an increased quantity of waste sludge.

A smaller but still significant cost factor is the power consumption which is a function of both the liquid to gas ratio required to remove SO_2 from the flue gas as well as the pressure required to make the gas flow through the system.

The design of the FGD system should provide for minimizing these costs under actual operating conditions. Prior to discussion of how these costs can be minimized it would be appropriate to consider the details of the Peabody FGD System shown in Fig. 1.

System Description

The Peabody FGD System is based upon the use of a patented, high-velocity spray tower as an absorber. This system offers the advantages of an open gas flow path which provides for high reliability because surfaces which scale or plug are minimized, low gas pressure drop which reduces system power requirements and adaptability to a wide variety of load conditions.

Fig. 1 shows the basic components of the Peabody FGD System. Flue gas enters the bottom of the absorber and is contacted countercurrently with a slurry containing a mixture of calcium sulfite, calcium sulfate and unreacted alkali. The slurry enters the tower via multiple spray headers. The number of spray headers is a function of both design and normal operating conditions. After leaving the absorption zone in the tower, where the gas has been contacted with the slurry, the scrubbed gas contains a significant amount of entrained slurry. The gas then flows upward through an interface tray and mist eliminator for removal of the entrained slurry prior to being discharged to the stack.

The slurry leaving the absorber flows by gravity to a slurry recycle tank where the reaction between SO_2 and the alkali goes to completion. The bulk of the slurry is then recycled to the absorber and excess waste slurry leaves the system as overflow from the slurry recycle tank.

A critical area in the design of the absorber is the interface tray. Entrainment of the slurry in the gas leaving the absorption zone is very significant. Operating data from the Detroit Edison St. Clair installation has shown, for example, that a 30 foot diameter absorber would have more than 100 gpm of slurry entrained in the gas stream leaving the absorption zone. The

alkali contained in the entrained slurry will react with the remaining SO_2 in the flue gas. Unless proper precautions are taken, a severe plugging problem at the interface tray due to calcium sulfite precipitation will occur. The method used to overcome this problem is to deluge the weeping tray with a washing medium and thus present a liquid barrier to the entrained slurry.

In many current designs the wash medium used is water, either fresh water, reclaimed water from the sludge dewatering system termed supernate, or a mixture of both; however, the quantity of wash water available is limited by the close loop requirement and maintenance of proper slurry concentrations within the absorber slurry system. Thus the quantity of wash water may be inadequate to eliminate plugging particularly at less than design load and percent sulfur in coal. Conversely, to adequately wash the tray at less than design conditions can result in an open loop water balance which is considered unacceptable. This is particularly true when scrubbing low sulfur coal flue gases.

A development made by Peabody, which presently is being patented, has been to utilize a hydroclone to provide the interface tray wash medium. Operating data has shown that the hydroclone, which is a liquid cyclonic classifier, can be used to classify the solids in the recycle slurry such that calcium carbonates can be separated from the calcium sulfite and calcium sulfate. Without calcium carbonate the slurry cannot react with SO_2 to create a plugging problem. The decarbonated slurry can then be used in whatever quantities are necessary for the interface tray washing without upsetting the water balance. Thus a closed loop water balance can be maintained under all operating conditions.

The hydroclone not only furnishes a means of providing a wash medium but can also be used to achieve a 100% utilization of alkali or a stoichiometry of 1.0. In the schematic shown the hydroclone would be sized to meet both the washing and stoichiometry improvement requirements. Overflow from the wash tank is routed to a baffled section near the overflow nozzle from the slurry recycle tank such that only decarbonated slurry leaves the system as

waste slurry. Carbonates removed from the slurry as hydroclone underflow would be routed back to the main stream of the slurry recycle tank where it becomes part of the slurry recycled to the absorber.

The waste slurry from the slurry recycle tank overflows by gravity to a waste slurry sump where it is pumped to a solids recovery system which could be a thickener and filter or a pond. In transporting slurry, maintaining a minimum velocity in the transfer line at all times, is mandatory to prevent settling of solids and eliminate plugging. Since the quantity of the waste slurry can vary widely with load conditions, it is not possible to maintain a non-plugging flow velocity in the transfer line unless other provisions are made. The slurry transfer system is operated on a constant velocity loop between the waste slurry sump and the solids recovery system. Supernate from the solids recovery system is added to the waste slurry sump in a quantity equal to the difference between the maximum and actual waste slurry quantity. This maintains a constant velocity in the slurry transfer line and thus eliminates plugging.

The basic design philosophy of Peabody's FGD System is that each component of the system should have its own function and that performance of one component should not depend on the other. Thus the absorber's only function is to remove SO_2 under varying operating conditions. The wash system does not depend on the absorber operating conditions. The hydroclone insures that alkali utilization does not change at different load conditions. The quantity of waste slurry produced does not affect the slurry transfer system.

Control Concept

System controls can be complex or they can be simple. The complexity of the control system can be adapted to reflect the philosophy of the owner as well as the requirement of the system.

Several years of experience with operating plants has shown that complex controls are not a requirement of the Peabody FGD System.

The same concept of designing an uncomplicated mechanical system also carries over into the control philosophy. Every attempt has been made to minimize controls wherever possible.

Using overflow from the wash tank and slurry recycle tank instead of a level controller is an example of this approach. The fewer the number of control loops, the lower is the risk of operating problems due to instrument malfunction.

Fig. 1 shows the basic controls of a Peabody FGD System.

Because the spray tower does not depend on gas velocity as a criteria of performance, gas flows from 0 to 100% of design can be accommodated. This means that flow control of the flue gas to each spray absorber, even in multiple absorber systems, is not necessary because of the adaptability of the spray tower to various gas flows.

SO_2 analyzers on the inlet and outlet gas streams provide the basis for establishing the required system performance. However, neither of these analyzers serve to control system variables. They serve only a monitoring purpose. The quality of the coal being burned establishes the allowable outlet ppm of SO_2 in the flue gas. This then becomes the guide to the operator for adjusting system operating variables to achieve desired performance and minimized operating costs.

pH control of the recycle slurry establishes the alkali makeup to the FGD System. As the pounds per hour of SO_2 to the system changes as a result of boiler load or sulfur content of the coal the pH controller correspondingly increases or decreases the amount of alkali to the system while maintaining the pH at the desired set point. Generally when multiple taps are made from an alkali slurry loop hydraulic conditions can vary. To compensate for these hydraulic variations, the pH controller resets a flow controller and thus insures a stable control condition.

Various ways can be used to control the quantity of fresh makeup water to maintain a closed loop water balance. A boiler load signal is but one of these ways.

To insure that the proper percent solids in the recycle slurry is maintained, to avoid scaling conditions a density control is used.

Supernate from the waste solids recovery system is returned to the slurry recycle tank by

means of a density controller. As the quantity of waste solids produced varies with operating conditions, the density controller modulates the quantity of supernate returned to the slurry recycle tank to maintain the proper percent solids.

During our Detroit Edison contractual performance run, high and low sulfur coals were burned interchangeably, with inlet SO_2 concentration cycling from 300 to 2700 ppm and load condition fluctuating between 30% to 100% of design. The control concept for Detroit Edison is the same as outlined above and has proved itself extremely adaptable to rapid changing operating conditions.

Operating Variables

Having discussed the components of the FGD System, their inter-relationships and the controls required to maintain desired performances, let us look into the operation of the system at other than design conditions.

Design conditions establish the quantity of flue gas to be scrubbed, the inlet SO_2 concentration as well as the pounds of SO_2 to be removed and the efficiency at which it has to be removed. Normal operating conditions generally result in a reduced quantity of flue gas, lower inlet SO_2 concentration, less pounds of SO_2 to be removed and most significantly a lower SO_2 removal efficiency.

The operator of the FGD System has basically two options open to him. The system can be operated as if design conditions prevailed at all times and no changes made to the system or the system variables can be adjusted to reflect ac-

Figure 2

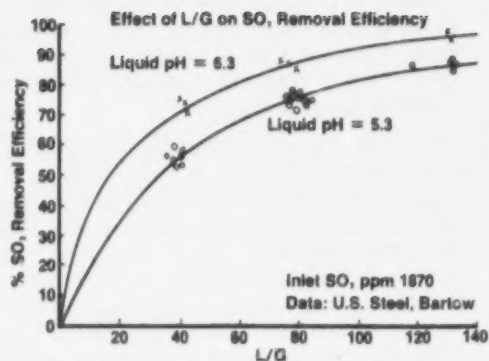
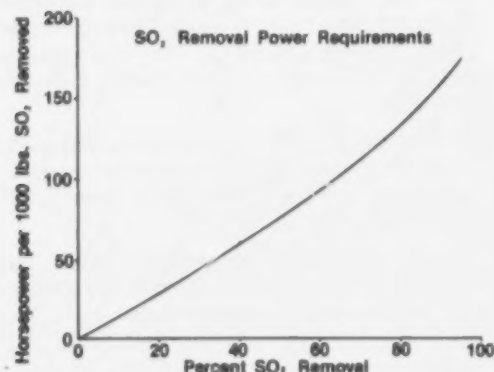


Figure 3

tual operating conditions in order to minimize operating costs. The operating variables available in order to reduce operating costs are:

- Quantity of gas scrubbed
- L/G ratio
- pH of recycle slurry

A. Quantity of Gas Scrubbed

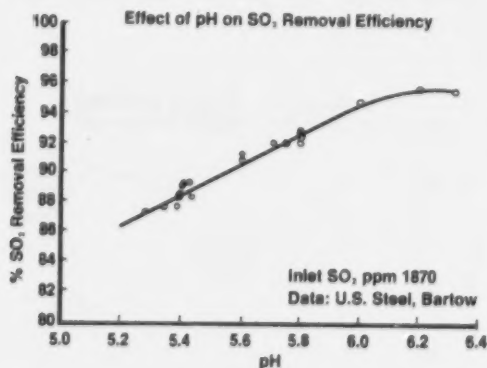
As SO_2 removal requirements decrease the tendency is to think that the most economical way to reduce operating cost is to by-pass as much of the gas as possible and scrub only a minimum portion of the gas. This is a fallacy. As more and more gas is by-passed around the system, the efficiency of removal of SO_2 in the portion left to be scrubbed increases significantly. The energy to remove SO_2 at a higher efficiency correspondingly increases the power requirements markedly. Fig. 2 shows the effect of removing SO_2 under different efficiency conditions. This curve is based upon an inlet SO_2 concentration of 2900 ppm using limestone as the alkali. As the SO_2 concentration decreases the power required to remove 1000 pounds of SO_2 will increase. The curve for other alkalis will have different horsepower values but the shape of the curve would be the same. From this curve it is seen that to remove a thousand pounds per hour of SO_2 at 90% efficiency would require 160 H.P.; however, if the same thousand pounds of SO_2 were to be removed at a 50% efficiency, the horsepower requirement is less than half or 75 H.P. The pounds per hour of SO_2 to be removed is the same regardless of how the FGD System is operated. Thus the most economical way of removing SO_2 is to scrub as much of the gas as possible which means that the SO_2 is removed at the lowest efficiency possible.

As a general rule, for FGD Systems consisting of multiple absorbers, an absorber should be shut down only when the remaining units have sufficient capacity to scrub all of the gas.

B. L/G Ratio

Fig. 3 shows plant operating data indicating the effect of the L/G ratio on SO_2 removal efficiency. As can be seen from these curves, as the required SO_2 removal efficiency decreases from the design point there is a marked reduction in the required L/G. For example, at a limestone slurry pH of 6.3 decreasing the required SO_2 removal efficiency from 90% to 70% decreases the required L/G ratio from 85 to 40. The net effect of this fact is that a power saving is possible when the L/G ratio is reduced to reflect the actual SO_2 removal efficiency requirements.

Figure 4



C. pH of Recycle Slurry

Control of the pH of the recycle slurry becomes another important factor in the control of SO_2 removal efficiency. As can be seen on Fig. 4, varying the pH of a limestone recycle slurry from 5.2 to 6.3 for a given L/G ratio causes a change in SO_2 removal efficiency from 86 to 96%.

Application of Operating Variables

The FGD Systems must be mechanically designed so that the previously described operating variables can be applied to minimize operating costs. This is achieved by several means:

a) Multiple Absorbers per FGD System

The number of absorbers per FGD System, Fig. 5, is dependent upon the design gas flow as well as the required flexibility. For example, a 600 MW system would require mechanically a mini-

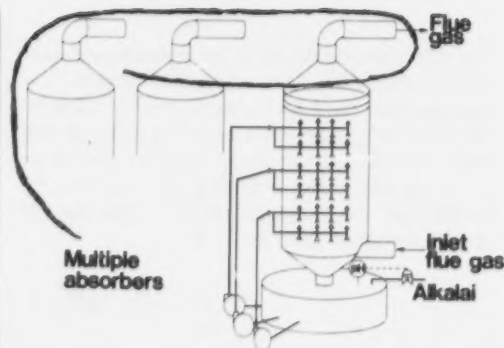


Figure 5

mum of 3 absorbers. However, if the power plant is to be operated at 75% load for long periods of time four smaller absorbers may be considered. The power saving resolution from maintaining one absorber off stream may justify the added capital expenses.

b) Multiple Slurry Recycle Pumps per Absorber

The minimum number of slurry recycle pumps per absorber, Fig. 6, is established by the total

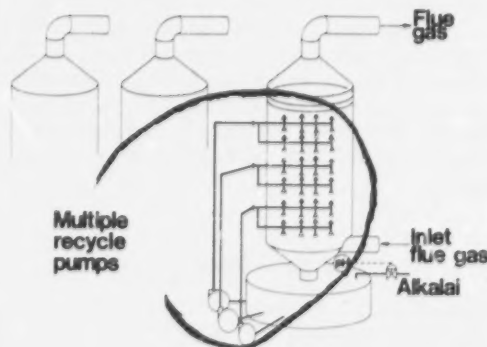


Figure 6

gallons of slurry being pumped. However, a greater number of smaller capacity pumps can be provided if additional flexibility is required. A typical system might have three recycle pumps each feeding a set of spray headers. No attempt is made to manifold one pumping system to the other. This improves reliability and allows each of the pumps to work independently of each other. If SO_2 removal efficiency has decreased such that L/G can be reduced, step wise control of the L/G is achieved by shutting off one or more of these pumps. Reducing the

L/G ratio also achieves a fan horsepower saving in addition to pumping horsepower. At design gas flow each slurry spray header contributes a $1\frac{1}{2}$ " w.c. pressure drop to the flue gas flow path. Thus where two spray headers are manifolded to one pump, reducing the L/G ratio by shutting off one pump per module reduces the system pressure drop by 1" w.c.

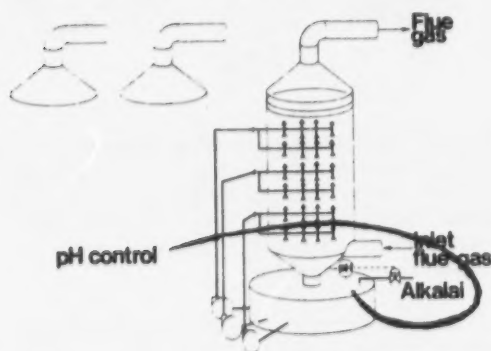


Figure 7 c) pH of Recycle Slurry

Step wise control of the L/G ratio produces a rough approach to the desired SO₂ removal efficiency. Control of the recycle slurry pH, Fig. 7, becomes the means of fine tuning the system. An operator need only to adjust the set point of the pH controller until the SO₂ analyzer meets the desired SO₂ concentration in the scrubbed gas.

Figure 8

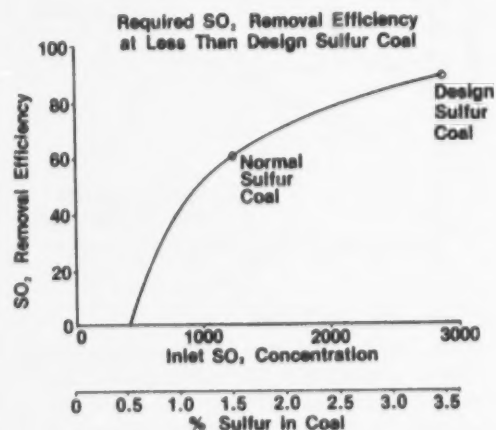
Typical Boiler Design Condition

Boiler Capacity — 600 MW
 Fuel — 3.5% S Design
 1.4% S Normal
 ACFM to Scrubber — 1,851,000
 Gas Inlet Temp. — 300°F
 Percent of Gas Scrubbed — 100%
 Gas Inlet SO₂ Concentration — Design 2900 ppm
 Normal 1170 ppm
 Allowable SO₂ Concentration in Stack — 450 ppm max.
 Required SO₂ Removal Efficiency — 89%
 Alkali — Limestone
 Design L/G — 90 Gal/1000 CFM
 No. of Spray Tower Modules — 3
 No. of Slurry Recycle Pumps per Module — 3

Example

Fig. 8 shows typical boiler design conditions for a 600 MW boiler illustrating some the principles discussed. Fig. 9 shows that, at the design sulfur coal, an 89% efficiency would be required but as the sulfur in the coal decreases, the required removal efficiency drops to slightly over 60% for the 1.4% sulfur coal. As a result of this lower efficiency requirement the number of pumping stages per module could correspondingly be decreased. Operating experience has shown (Fig. 10) that each of the three pumps operates as a stage with an equal SO₂ removal efficiency. When one pump per absorber operates, an SO₂ removal efficiency of 52% is achieved. With two pumps per absorber operating, a 78% removal efficiency is achieved. Three pumps in operation provides the design SO₂ removal efficiency of 89%.

Figure 9



For the normal 1.4% sulfur coal condition where a 60% removal efficiency is required, operating with one pump per absorber at a 52% efficiency would be inadequate, two pumps per absorber produces a 78% efficiency which is too high. The two pumps per absorber method of operation would be selected and the pH of the recycle slurry reduced until the required 60% efficiency, as indicated by the 450 ppm SO₂ concentration in the stack gas, is reached. At this condition the amount of SO₂ removed and the amount of sludge produced is no more, nor no less than required to meet emission standards.

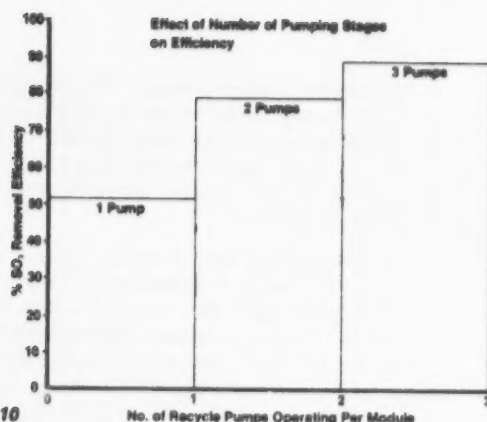


Figure 10

Fig. 11 shows the potential power saving possibilities under different load and inlet SO₂ concentrations. Zero to 33% of design gas flow requires one module on stream. 33 to 67% of design gas flow requires two modules on stream. 67 to 100% of design gas flow requires 3 modules on stream. With inlet SO₂ concentrations of 450 to 850 ppm a minimum of one recycle pump per module is required. Inlet concentrations of 850 to 1900 ppm requires a minimum of two recycle pumps per module. 1900 to 2900 ppm of SO₂ requires all three recycle pumps per module operating. For each of these variations in operating conditions, the absorber pressure drop (D.P.) and power savings are shown (P.S.). At design gas flow with an inlet SO₂ concentra-

tion of 1170 which corresponds to the normal 1.4% sulfur coal condition a power saving of 2466 H.P. is achieved. This is a 33% saving in horsepower.

However, the potential horsepower savings and benefits do not end here. For illustrative purposes a gas flow 80% of the design value is used. From the systems which Peabody has designed and quoted this flow rate realistically approaches an actual full load condition. The balance 20% usually represent design safety factors.

Fig. 12 shows the alternate methods of operation possible at a gas flow 80% of design under different SO₂ concentration conditions.

For the case where three absorbers are on stream, with the gas flow to the absorbers decreased the actual L/G ratio increases, in this case by 25%. The result is a system which has a SO₂ removal efficiency capability greater than the design 89%. This permits shutting off a recycle pump at a higher inlet SO₂ concentration than indicated earlier, in this case 2150 ppm.

As the inlet SO₂ concentration decreases, by-passing some of the gas and shutting down absorbers while maintaining 450 ppm SO₂ in the stack gas is a possible method of operation. At approximately 1800 ppm inlet SO₂ concentration it is possible to shut down one absorber, thus making two alternate methods of operation possible. At 750 ppm inlet SO₂ concentration it

Figure 11

**Power Savings Under Reduced Load Conditions
(Basis: 100% Gas Scrubbed)**

% of Design Gas Flow	Inlet So ₂ Conc		
	450 - 850 ppm	850 - 1900 ppm	1900 - 2900 ppm
0 - 33	1 Module 1 Pump/Module D.P. = 1½" W.C. P.S. = 6258 H.P.	1 Module 2 Pumps/Module D.P. = 2½" W.C. P.S. = 5118 H.P.	1 Module 3 Pumps/Module D.P. = 3½" W.C. P.S. = 3978 H.P.
33 - 67	2 Modules 1 Pump/Module D.P. = 1½" W.C. P.S. = 5595 H.P.	2 Modules 2 Pumps/Module D.P. = 2½" W.C. P.S. = 3792 H.P.	2 Modules 3 Pumps/Module D.P. = 3½" W.C. P.S. = 1989 H.P.
67 - 100	3 Modules 1 Pump/Module D.P. = 1" W.C. P.S. = 4932 H.P.	3 Modules 2 Pumps/Module D.P. = 2½" W.C. P.S. = 2466 H.P.	3 Modules 3 Pumps/Module D.P. = 3½" W.C. P.S. = 0

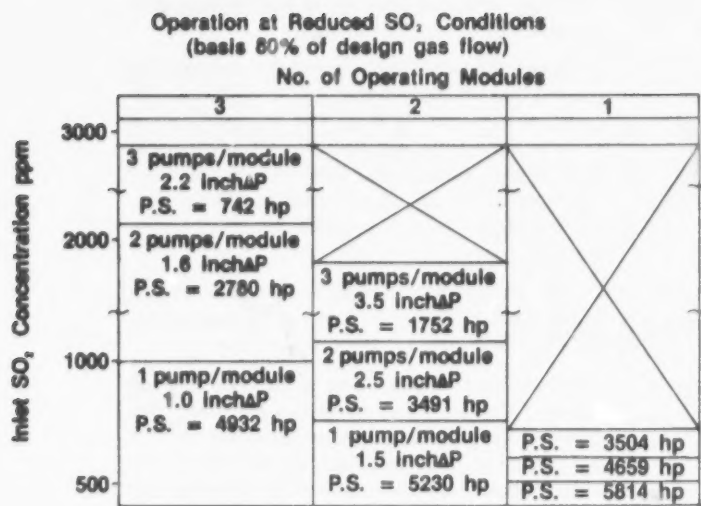


Figure 12 is possible to shutdown two absorbers making three methods of operation possible.

The power saving achieved for the alternate methods of operation resulting from shutting down absorbers and recycle pumps are shown in Fig. 12.

At the expected normal inlet SO₂ concentration of 1170 ppm corresponding to the 1.4% sulfur coal, it appears that the greatest power savings is achieved with two modules on stream and by passing some of the gas. The earlier generalization that scrubbing all of the gas minimizes power consumption still holds true. However, because the L/G ratio is controlled in a stepwise manner exceptions do occur. This is one of them.

By looking at this chart another significant factor emerges. In designing the system for design conditions no consideration had been given to a spare module. However, for inlet SO₂ concentrations of 1800 ppm equivalent to 2.2% sulfur coal or less where two or more alternate methods of operation are possible a spare module

does exist. Thus the owner of this system would have the capability of handling peak operating condition should they occur and yet for no additional capital cost have the added security of a spare module for a wide range of normal operating conditions.

Had no attempt been made to apply the principles used for 80% of design gas flow and the 1.4% sulfur coal, SO₂ removal would have exceeded the recommended requirements by 5200 lbs./hr. of SO₂. At \$8 a ton for limestone and \$10 a dry ton for waste sludge this would represent an added operating cost of \$850,000 per year. This cost saving in addition to the ability to save approximately 50% of the power requirement or 3491 H.P., highlights the significance of applying the principles outlined.

Summary

a) Alkali consumption and sludge handling costs are minimized by adjusting the L/G ratio and slurry pH to insure that no more than the required amount of SO₂ is removed.

b) The use of the hydroclone provides the means for achieving 100% alkali utilization and thus stoichiometry is no longer an operating cost factor.

c) Pumping power can be reduced by shutting off absorbers and reducing the L/G ratio to reflect load and SO₂ removal efficiency requirements.

d) Operating the absorbers at reduced L/G ratio by shutting off one or more recycle pumps per absorber decreases the system pressure drop with corresponding fan horsepower saving.

In summary it can be stated that the Peabody FGD System is extremely adaptable to a wide variety of operating conditions. What becomes significant in designing these systems is not only an understanding of the design conditions but also the normal operating conditions. This permits designing a system which produces the lowest operating cost to the owner.

BY-PRODUCT-UTILIZATION/ULTIMATE-
DISPOSAL OF GAS CLEANING WASTES FROM
COAL-FIRED POWER GENERATION

by

William Ellison^{1/} and Edward Shapiro^{2/}

ABSTRACT

A review is given of pollution control rules in solids disposal called for by Public Law 94-580, The Resource Conservation and Recovery Act of 1976, and now formally proposed by EPA. The text of these requirements indicates that the disposal of solid collected and formed in cleaning of flue gas from coal-fired boilers can be expected to be subject to comprehensive environmental regulations governing design, permitting and operation of storage and disposal sites. Levels of concentration of heavy trace elements in fired coal will directly influence hazardousness in disposal of both fly ash and scrubber sludge, as well as the complexity of controls required for groundwater protection. A review of recently published details of EPA stir-testing procedures and toxicity data on mixtures of wet-collected fly ash and scrubber sludge indicates that fly ash and SO₂ solids may be expected to regularly test out as non-hazardous. At the same time, isolated disposal and broad availability of current and future fly ash production for by-product utilization may be significantly limited. This is the result of anticipated increasing use of dry fly ash catch as a fixation reagent to stabilize SO₂ sludge for improved ultimate disposal. Major site-related hydrological and geological factors affecting selection, design and operation of sites for permanent disposal of raw and stabilized solid wastes are reviewed along with techniques for preliminary investigation of acceptability of existing facilities and available sites. Utilization of fly ash and pyrites wastes in flue gas desulfurization (FGD), and emerging of commercial design technology for broadened application of coal-based regenerative FGD technology is described along with interrelated progress in management of coal cleaning waste.

^{1/}Assistant General Manager, Air Pollution Control, NUS Corporation,
Rockville, Maryland

^{2/}Vice President, Engineering, Pittsburgh Environmental and Energy
Systems Incorporated, Pleasant Hills, Pennsylvania

INTRODUCTION

FGD sludge and fly ash from cleaning of coal-fired boiler flue gas are a major waste product of the utility industry. These materials are of concern because of the large quantities generated as well as the possible adverse pollution effects due to the waste properties. Because of the hydrological and geological complexity of typical solid disposal storage sites such as illustrated in Figure 1, many environmental questions, including that of potential long-term effects on groundwater quality, can arise at a major facility. Thus, fly ash and scrubber sludge management currently poses a significant technological, environmental and transportation problem.

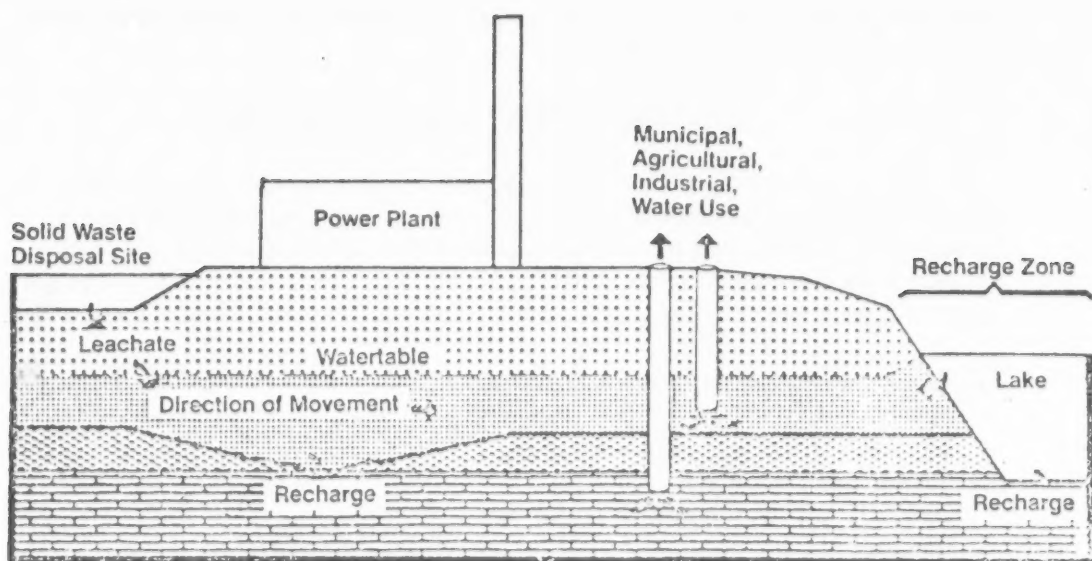
In recognition of potential effects on human health and the environment resulting from the improper disposal of solid wastes, the Solid Waste Disposal Act, as amended by the Resource Conservation and Recovery Act of 1976 (Public Law 94-580^{3/}), has resulted in comprehensive performance standards^{4/} and, in the case of wastes determined to be hazardous, has established a management control system requiring "cradle-to-grave" cognizance, including appropriate monitoring, record-keeping and reporting throughout the system^{5/}. Planning and design of waste disposal facilities is thereby a major engineering undertaking requiring effective integration in design and operation of pollution control and ultimate disposal systems after thorough geotechnical and hydrological study of the site^{6/}.

^{3/}Public Law 94-580, 94th Congress, Resource Conservation and Recovery Act of 1976, October 21, 1976.

^{4/}U.S. Environmental Protection Agency, "Solid Waste Disposal Facilities, Proposed Classification Criteria", 40 CFR-257, Federal Register, Vol 43, No. 25, pp 4952-4955, February 6, 1978.

^{5/}U.S. Environmental Protection Agency, "Hazardous Waste, Proposed Guidelines and Regulations and Proposal on Identification and Listing" 40 CFR-250, Federal Register, Vol. 43, No. 243, pp 58946-59028, December 18, 1978.

^{6/}Ellison, W. and R.S. Kaufmann, "Toward Safe Scrubber-Sludge Disposal", Power, pp 54-57, July, 1978.



Key

- | | |
|---|--|
|  Zone of Aeration |  Confining Layer |
|  Unconfined Aquifer (Zone of Saturation) |  Confined Aquifer |
| |  Lake |

FIGURE 1
Interaction of Solid Waste Disposal
and Groundwater Systems

SOLID WASTE MANAGEMENT TO MEET RCRA/EPA REGULATIONS

A review of pollution control guidelines for ultimate disposal of solid waste called for by Public Law 94-580, The Resource Conservation and Recovery Act of 1976, and now issued by EPA as proposed rules indicates that the disposal of boiler flue-gas-scrubber solid wastes from flue gas desulfurization (FGD) as well as collected fly ash can be expected to be subject to comprehensive controls and standards under new governmental regulations.

RCRA/EPA Criteria for Disposal of Solids

While these general provisions, which are made under Subtitle D, do not contain as stringent operating and monitoring requirements as apply to hazardous waste disposal facilities under Subtitle C, they impose a comprehensive new discipline for all large-scale storage or disposal of solids. Major aspects of the criteria are as follows:

1. Broad provisions to prevent siting of disposal facilities in environmentally sensitive areas including wetlands, floodplains, sole source aquifers, etc.
2. Protection of surface water bodies through extension of the NPDES permitting requirement for any point source discharge and by control of non-point sources to prevent or minimize pollutant discharges
3. Stringent provisions as follows protecting groundwater in usable aquifers, i.e. those containing less than 10,000 mg/l total dissolved solids, (except those that have received State designation for a use other than as a drinking water supply for human consumption):
 - (a) Quality of groundwater beyond the disposal-facility property-boundary shall not be endangered by the facility, i.e. degraded such that more extensive treatment is thereby required to prepare the water for drinking water purposes.
 - (b) Prevention of endangerment shall be assured either by the collection and proper treatment and disposal of leachate produced by the facility or alternatively, by site selection and facility design to adequately control migration of leachate from the disposal structure.
 - (c) Where appropriate, prevention of endangerment shall be verified by a suitable groundwater monitoring program.

Evaluation of Solids Disposal Sites and Facilities

EPA Inventorying of Deficient Sites

Subtitle D also requires that EPA immediately proceed to compile a nationwide inventory of objectionable solid waste disposal sites. Using readily available geologic and hydrologic data, all existing facilities are to be assessed by the government for potential contamination of groundwater and surface water so that necessary in-depth site evaluations including field collection of detailed data can be selectively carried out where necessary. EPA's new SIA (surface impoundment assessment) evaluation system⁷ which yields a first-round numerical approximation of the relative environmental impact of waste impoundments, provides means for consistency in this preliminary prioritizing assessment of existing facilities.

Applicability of SIA System

The SIA system consists of a step-by-step procedure for collecting available site information from which the water-pollution potential of a specific facility may be estimated. SIA can also be used to calculate the comparative pollution potential of each of several alternative sites being considered for future use in solids storage or disposal so as to provide an initial ranking of site acceptability.

Parameters Influencing SIA Score

The SIA system makes a best available estimate of the following principal parameters bearing on the two major concerns in control of water pollution from disposal of solids:

Groundwater Contamination

The potential for groundwater contamination is most strongly influenced by:

- (1) The permeability of the existing earth material and the thickness of the unsaturated zone above the groundwater, which tends to protect the underlying aquifer
- (2) The permeability of the existing earth material and the thickness of the saturated zone (the aquifer itself) which represents the quantity of groundwater availability from the aquifer
- (3) The usefulness of this groundwater as determined by whether the aquifer is currently in use as a source of drinking water, or if not, the total dissolved solids (TDS) content of the groundwater

⁷/U.S. Environmental Protection Agency, "A Manual for Evaluating Contamination Potential of Surface Impoundments", Report No. 570/9-78-003, June, 1978.

- (4) The potential for presence of critical contaminants in the waste as judged by the waste-material source or type and the corresponding EPA hazard potential rating .

Endangerment of Current Water Supplies

The potential endangerment of water supplies currently in use is most strongly influenced by:

- (1) The degree of vulnerability to contamination by the impoundment as judged by the type of existing water use, i.e. water well or surface supply
- (2) The degree of anticipated attenuation of contaminant flow as related to the distance of existing water uses from the impoundment
- (3) The proportion of contamination flow that may reach use-points as determined by the location of existing water uses with respect to the anticipated flow direction of the contaminated groundwater .

Interpretation of SIA Scores

EPA's SIA manual includes charts tabulating selected values of each of these parameters and displaying corresponding standardized numerical scores. The estimate of contamination potential is obtained by summing individual score components for its parameters. This total groundwater-contamination-potential score can range from 1 to 30. In comparing sites and making a preliminary site evaluation, a score of 10 or less may mean a potentially good site for disposal while a score of 20 or more indicates that there may be significant problems. The supply-endangerment score will range between 0 and 9, 4 or less being favorable. Use of the SIA system makes it possible to identify specific low-scoring sites and thereby justify more detailed study of them for the purpose of verifying future use for disposal; or, in the case of existing facilities, to identify high-scoring sites and thereby justify further investigation of them to ascertain possible need for remedial measures.

Utilization of SIA System by Owners of Facilities

The SIA system and the EPA manual in which it is presented affords the owner's staff the opportunity to gain valuable technical understanding of water pollution control effects of solid waste disposal and appropriate design and operating practices. Its timely use can help guide assembly of site information in a standardized format that can be understood by all concerned individuals while at the same time providing an early indication of possible actions required to achieve compliance with RCRA-EPA rules.

It should be noted however that the SIA evaluation will be made most effectively and lead to most meaningful scoring if it is carried out by a professional geologist who is capable of adequately characterizing geologic environments and of selecting suitably conservative values for the applicable parameters.

Requirements for Design and Operation of Hazardous-Waste Disposal Sites

Section 3004 of RCRA Subtitle C addresses standards applicable to owners and operators of hazardous-waste storage, treatment, and disposal facilities. These regulations, now issued by EPA as proposed rules, define the levels of environmental protection to be achieved and provide the criteria against which EPA will assess applications for permits to design and operate disposal and storage sites for hazardous waste.

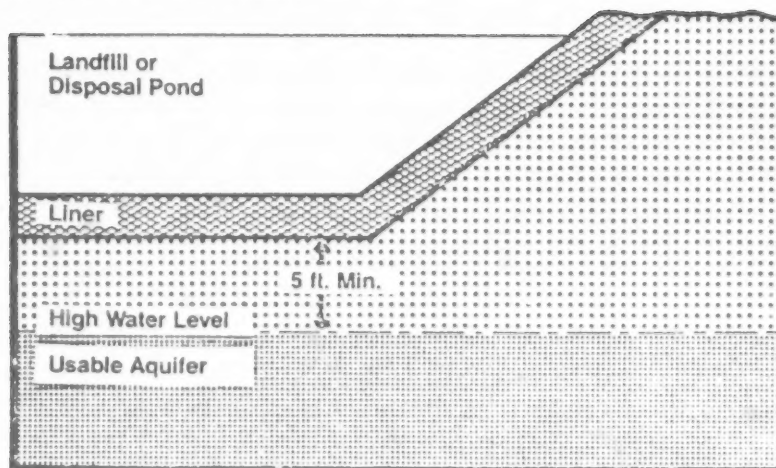
Of particular significance are RCRA-EPA-prescribed alternative leachate-containment methods applicable to hazardous waste facilities located above usable aquifers, (containing less than 10,000 mg/liter total dissolved solids). See Figure 2-A, B, and C:

Landfill Containment

Landfills over usable aquifers must be either:

- (a) Arranged as per Figure 2-A, when natural geologic and climate conditions allow, to provide a containment in the form of a natural liner of soil meeting specified criteria and at least the equivalent of a 10-foot thickness with a permeability no greater than 1×10^{-7} cm/sec (but with thickness no less than 5 feet), or
- (b) Constructed and operated such that leachate formed can be contained and removed from the landfill site when natural conditions do not permit the above containment. Under this option, per Figure 2-B, a limited containment shall be provided consisting of a minimum 5-foot thickness with a permeability no greater than 1×10^{-7} cm/sec. This liner shall be sloped at a 1% minimum grade so that the leachate is drained directly by gravity to a collection sump for removal and be overlain with an implied permeable layer of material such as gravel or sand so that any generated leachate can move rapidly to the sump. (Alternatively, the regulation allows a double liner and leachate-collection installation, the upper liner being a soil liner at least 3 feet thick with permeability no greater than 1×10^{-7} cm/sec.)

A. Natural Containment



B. Leachate Removal (Landfill)

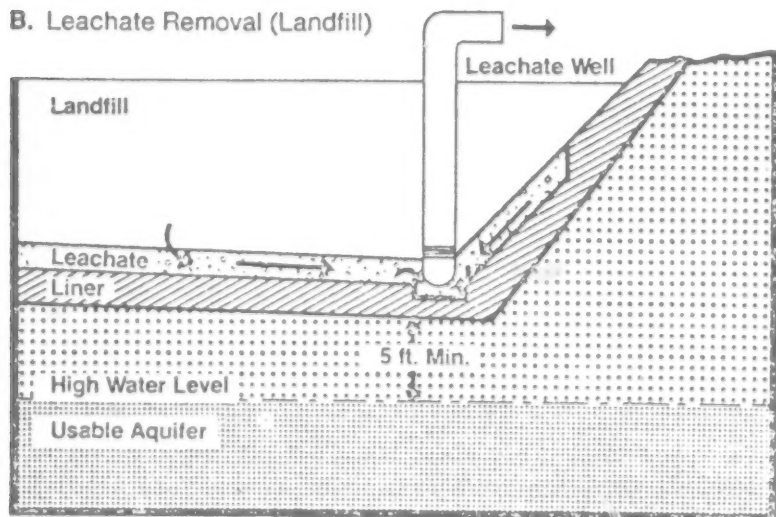


FIGURE 2 RCRA-EPA-PREScribed ALTERNATIVE
GROUNDWATER PROTECTION METHODS

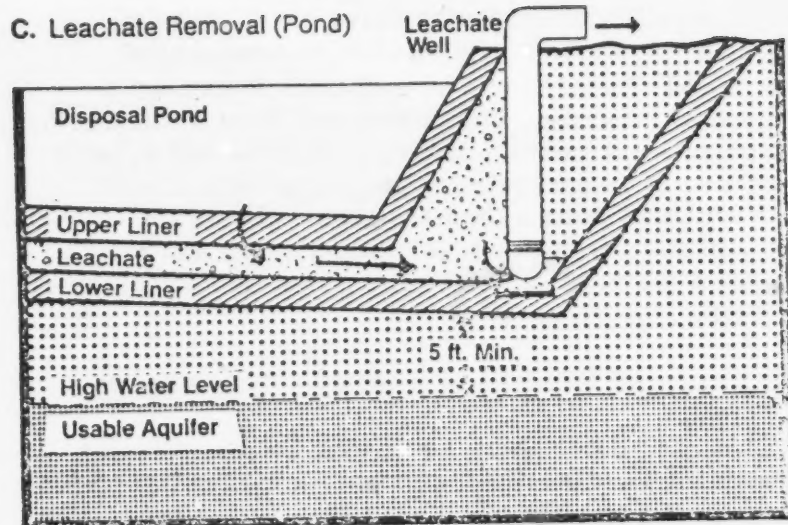


Fig. 2C.
RCRA—EPA—Prescribed
Alternative Groundwater
Protection Methods

Pond Containment

Disposal-pond type impoundments over usable aquifers must be either:

- (a) Arranged as per Figure 2-A, when the conditions allow, to provide a containment along the bottom and sides of the pond in the form of a natural liner of soil meeting specified criteria and at least the equivalent of a 10-foot thickness with a permeability no greater than 1×10^{-7} cm/sec, (but no less than 5 feet thick and no more than 1×10^{-7} cm/sec permeability.), or
- (b) Constructed and operated such that leachate formed can be contained and detected between a top liner and a bottom liner and removed when the conditions do not permit the containment above. Under this option, per Figure 2-C, the top liner shall be constructed of specific reconstituted clays or of artificial materials meeting prescribed criteria with permeability no greater than 1×10^{-7} cm/sec and of sufficient thickness to ensure mechanical integrity. The bottom soil liner shall be of natural in-place soil meeting specified criteria and at least 5-foot thickness with a permeability no greater than 1×10^{-7} cm/sec. (An artificial bottom liner may be used only for temporary disposal sites.) The leachate detection and removal system is to be a gravity flow drainage system installed between the top and bottom liners.

Assessment of Criteria for Solids Hazardousness Identification

Section 3001 of RCRA Subtitle C includes criteria for ascertaining hazardousness based on testing of a representative sample of the solids source. In view of the possible presence of significant trace quantities of heavy metals originating from the fired coal, toxicity evaluation is a key factor in determining if FGD waste and fly ash from a specific source is hazardous.

Solids Evaluation by Toxicant Extraction Procedure (TEP)

Provisions for site-specific examination for toxicity anticipate testing the potency of the leachate (extract) yielded by the proposed EPA Toxicant Extraction Procedure (TEP) applied to a sample of the actual solids. The proposed TEP calls for dewatering the representative solid waste sample, (crushing it if necessary to pass a 3/8" standard sieve), and stirring it in contact with 16 times its weight of deionized water while controlling

pH at 5.0 by addition of acetic acid during a 24-hour extraction period. A minor adjustment in the volume of the separated liquid extract, (by addition of deionized water), and the combining of this liquid mixture with the liquid-phase material formed in the dewatering of the sample results in a final liquid product, referred to as the TEP extract.

EPA's proposed rules indicate that in application of the TEP a solid waste will be deemed to be toxic if the TEP extract contains a concentration of any of the trace elements listed in EPA's National Interim Primary Drinking Water Regulation, (NIPDWR), greater than or equal to ten times the concentration allowed in drinking water by NIPDWR. (See Figure 3 which identifies the potentially toxic constituents listed in NIPDWR as well as the NIPDWR-allowed concentrations of these substances in drinking water and the ten-fold higher critical/threshold concentrations for TEP extract corresponding to a critical toxicity level in the tested solids.) This factor of ten reflects EPA's judgment that underground strata will attenuate inorganic concentrations in leachates ten-fold. Thus a leachate potency corresponding to the critical extract strength, i.e. containing one or more of the trace elements at a concentration of 10 X NIPDWR, is expected to be gradually mitigated, the leachate ultimately entering the groundwater system with concentrations of trace elements no greater than NIPDWR.

Generic Assessment of Typical Fly-Ash-Containing FGD Sludge

The TEP results in a twenty-fold mixing/dilution effect (weight ratio of water feed to sludge sample feed) by the use of extraneous water in forming the TEP extract. Therefore if it were assumed that:

- (a) all inorganic substances in the extract originate solely from liquid-phase material in the dewatered sludge sample and from the liquid-phase material formed in the dewatering of the sludge sample and
- (b) the free-moisture (raw-occluded-liquor) content of sludge is 50% by weight as in typical plants

then the TEP extract concentrations would correspond to occluded-raw-liquor concentrations that are $20 \div 0.50 =$ forty-fold higher than those in the extract. Since a significant amount of the trace element content of the extract may originate from the solid-phase material in the sample, the proportionality between the liquor and extract concentrations could be expected to be substantially lower than 40/1 in actual sludge-sample testing, e.g. in the range of 10/1 to 20/1. This reduction of concentration in the extract as compared to that in the waste, reflected in TEP, corresponds to the diluting effect of percolating rainwater that causes

FIGURE 3
CRITICAL CONTAMINANT CONCENTRATIONS
IN TEP EXTRACT FOR DESIGNATION
OF WASTES AS HAZARDOUS

Contaminant	Drinking Water NIPDWR Level, (EPA Primary Std) Milligrams per liter	Critical Extract Level, Milligrams per Liter (10 X NIPDWR)
Arsenic	0.05	0.50
Barium	1.	10.
Cadmium	0.010	0.10
Chromium	0.05	0.50
Lead	0.05	0.50
Mercury	0.002	0.02
Selenium	0.01	0.10
Silver	0.05	0.50
Endrin (1,2,3,4,10,10-hexachloro-6 7-epoxy-1,4,4a,5,6,7,8,8a-octahydro-1, 4-endo, endo-5, 8-dimethano naphthalene).	.0002	0.002
Lindane (1,2,3,4,5,6-hexachlorocyclohexane gamma isomer).	0.004	0.040
Methoxychlor 1,1,1-Trichloroethane) 2,2-bis (p-methoxyphenyl).	0.1	1.0
Toxaphene (C ₁₀ H ₁₀ Cl ₈ -technical chlorinated camphene, 67-69 percent chlorine).	0.005	0.050
2,4-D, (2,4-Dichlorophenoxyacetic acid).	0.1	1.0
2,4,5-TP Silvex (2,4,5-Trichloro- phenoxypropionic acid).	0.01	0.10

the leaching phenomenon. Thus sludge liquor concentrations that are capable of yielding critical extract concentrations will be typically greater than ten times ten X NIPDWR or 100 X NIPDWR.

On the other hand, recent publication^{8/} of results of sludge management study activity by EPA centered on disposal tests at TVA's Shawnee Station, Paducah, Kentucky, includes extensive field data on trace-element chemical-composition of this surface moisture in fly-ash-containing limestone type FGD systems at bituminous and sub-bituminous coal fired utility plants, (with wide-ranging proportions of fly ash), which indicate comparatively low concentrations. Ratioed to NIPDWR this data indicates that inorganic constituents affecting drinking water quality are present typically at concentrations only five times greater than NIPDWR and at levels no greater than twenty times NIPDWR. Thus fly-ash and scrubber sludge can be expected to be typically found to be non-hazardous as a result of TEP measurement by a margin of over 5/1 and to contain significantly less than the EPA-defined critical/threshold concentration of leachable toxic inorganic contaminants.

ASH/COAL UTILIZATION IN FGD SYSTEM DESIGN

Coal-based energy generation and attendant production of fly ash provide significant opportunity to gain maximum benefit of coal and fly ash use.

Fly Ash as Fixation Agent in FGD Sludge Stabilization

Means of Stabilizing Gas Cleaning Waste

Sludge stabilization is chemical processing to fix scrubber sludge so as to facilitate improved handling, transportation, placement, and consolidation at the ultimate disposal site. A common method now used at new coal-fired units includes pug mill blending of mixtures of dry fly ash with FGD filter cake and other dry additives followed by several days curing before compaction at a landfill disposal site.

^{8/}Rossoff, Jerome, et al., The Aerospace Corporation, "Landfill and Ponding Concepts for FGD Sludge Disposal", U.S. Environmental Protection Agency, Industry Briefing Conference on Technology for Lime/Limestone Wet Scrubbing, Research Triangle Park, North Carolina, p. 4, August 29, 1978.

Benefits of Solids Stabilization

Extensive test program work by EPA to date verifies that this method of chemical treatment significantly improves the load bearing characteristics of FGD sludge, decreasing the solubility of the major chemical species by a factor of two to four, and reducing sludge permeability by an order of magnitude^{9/}. Thus stabilization of FGD sludge, made possible by the presence of substantial quantities of fly ash available at coal-fired plants can help to insure effective ultimate disposal at a landfill site. In some instances, land may be improved and be reclaimed for beneficial use. At the same time, stabilization acts to seal in the surface moisture occluded in the sludge (which serves as a vital purge of soluble solids) reducing leachate formation and recycle to the FGD system.

Dry Scrubbing Waste

Dry-scrubbing type FGD systems have been extensively demonstrated during 1978 in lignite-fired boiler service and a number have been ordered during 1978 for full-scale application in bituminous, sub-bituminous and lignite service. This new FGD method uses slaked lime in a spray-type gas absorber comparable to a spray dryer to collect SO_2 and convert it to a dry reaction product. The desulfurized flue gas flows to a fabric filter or electrostatic precipitator to collect dry SO_2 -reaction-product and fly ash. The waste product from dry scrubbing is a dry mixture of fly ash, unused lime and calcium sulfite and sulfate comparable to stabilization mixtures from wet-scrubber type FGD systems and thus may be expected to be a self-stabilizer in landfill disposal.

Fly Ash as FGD Reagent

Background

Extensive testing and application of wet scrubbers on lignite and sub-bituminous service has demonstrated the effectiveness of alkaline fly ash in absorbing flue-gas SO_2 . In conjunction with lime or limestone addition, SO_2 removal efficiency ranges as high as 65 to 90%. Commercial scrubber installations designed to utilize fly ash to achieve FGD include Montana Power Company's Colstrip Station Unit Nos. 1 and 2, Northern States Power Company's Sherburne County Station Unit No. 1, Minnkota Power Cooperative's M. R. Young Station Unit No. 2, and Minnesota Power and Light Company's Clay Boswell Station Unit No. 4.

^{9/}U.S. Environmental Protection Agency, "Control of Waste and Water Pollution from Power Plant Flue Gas Cleaning Systems: First Annual R and D Report", Report No. EPA-600/7-76-018, p. 2, October 1976.

SO₂ Absorption and Gypsum Scale Control

In recirculating scrubbing slurries containing a substantial level of alkaline western fly ash, the dissolving of alkali metal components of the fly ash results in a buildup of soluble sulfites, principally sodium and magnesium, in the liquid phase. This dissolved active alkali helps achieve a high SO₂ removal rate, forming bisulfite ions, which react with lime, the predominant fly-ash alkali component, to precipitate the SO₂ catch as calcium sulfite and sulfate. The elevated sulfite ion concentration may depress the dissolved calcium content, (because of the very low solubility of calcium sulfite), to a level sufficient to reduce the calcium and sulfate to an unsaturated level. In this gypsum-scale-controlled mode, sulfate precipitates from the system by attaching itself as calcium sulfate to the principal calcium-sulfite precipitate-crystal, forming a co-precipitate. Recent disclosure of tests on full-scale scrubbers at Arizona Public Service Company's Four Corners Station Unit Nos. 1 and 2,^{10/} supplemented by electron micrographs indicates that at fly-ash slurry solids concentration in excess of 5%, calcium sulfate precipitate from the unsaturated liquid phase nucleates on the fly ash particles. Due to absence of abrasive effects at elevated fly-ash slurry solids concentrations, APS has concluded that precipitation of SO₂ catch on the fly ash particles significantly reduces the abrasiveness of the suspended fly ash particles.

Regenerative FGD Selection

General

Appropriate selection of by-product FGD technology to suit site conditions offers a major degree of flexibility in meeting environmental pollution control requirements. With such systems collected fly ash emission can be fully isolated from other waste materials for immediate or future use in meeting growing market demands for this by-product material. Regenerative-type SO₂ removal systems provide a basis for meeting the sludge disposal problem by converting the bulk of the SO₂ catch to saleable by-product: gypsum, sulfuric acid, elemental sulfur or other marketable products. Although limited by the size and locality of gypsum markets, by-product gypsum manufacture can offer important economies in view of the simplicity of FGD system design in this regenerative process mode. Significant advances have been made in West Germany in the last few years in application of regenerative FGD systems to yield commercial gypsum by-product.

^{10/}Nelms, W. M., and C. F. Turton, "Sulfur Removal Testing of Particulate Scrubbers at Four Corners". Arizona Section of American Society of Mechanical Engineers, Phoenix, Arizona, pp 13-15, January 12, 1978.

Waste Management

Fly-ash collection is carried out upstream of FGD, preferably by dry collectors, thus permitting fly-ash to be stored in a landfill-type facility without generation of process liquid effluents other than leachates produced and controlled at the disposal site. The FGD system, generally of the wet-process type, will generate a process effluent in the form of a purge of scrubbing liquid containing non-precipitating components including chlorides originating in the fired coal. Minimum requirements for liquid effluent treatment prior to discharge typically include reaction with limestone or slaked lime at elevated pH to precipitate heavy metal components.

Utilization of Pyrites Waste and Coal for Reagent/Energy Needs

The State of Pennsylvania has recently announced current full-scale installation for 1979 startup at a small State-owned coal-fired power plant in Pittsburgh of an FGD system that can utilize ferrous sulfide from pyrites waste as the chemical absorption reagent. This gas cleaning technology of Pennsylvania Environmental and Energy Systems, Incorporated, (PENSYS), Pittsburgh, Pennsylvania, called Sulf-X Process, is based on use of waste iron feedstocks including concentrated pyrites waste from bituminous coal cleaning or non-ferrous mining and milling to make up system iron losses. The unique chemical characteristics of ferrous sulfide maintains an absorption-reduction process mode resulting in efficient simultaneous removal of SO_2 and NO (nitric oxide). Common coal may be combusted to calcine the pyrites makeup and the spent ferrous/sulfur chemical thereby producing elemental sulfur by-product in vapor form and regenerating the ferrous sulfide reagent for reuse in the absorption-reduction gas cleaning step. Development of this pyrites-fed regenerative FGD process for commercial service has been under way since 1976 through coal-based experimental and demonstration programs including:

- (a) Wet scrubber pilot plant gas cleaning operation and testing for the Department of Army, the Pennsylvania Science and Engineering Foundation of the Pennsylvania Department of Commerce, and the Appalachian Regional Commission.
- (b) Testing and assessment of technology for gas cleaning, sulfide regeneration, and sulfur by-product manufacture for the Appalachian Regional Commission

- (c) Process design evaluation through laboratory, bench scale and pilot/demonstration stage testing for the U.S. Department of Energy

CONCLUSIONS

Current energy planning based on increased use of coal-fired boiler plants in the United States requires support through continued development and availability of improved techniques for controlling environmental impact. Commercial by-product utilization and/or acceptable ultimate disposal of by-product solids generated by stack-gas cleaning systems at existing and new utility and industrial plants is of major significance in limiting environmental pollution.

Moreover, compliance with new RCRA/EPA solid waste regulations applicable to storage and/or disposal of scrubber sludge and collected fly ash will be most economically achieved by integration, both in design and operation, of pollution control and waste handling facilities, and by adequate evaluation of site conditions so as to achieve advantageous production and utilization of marketable by-product forms where applicable and to assure effective containment of contaminants in solids which are to be discarded or stored.

**FLUE GAS DESULFURIZATION AND FERTILIZER MANUFACTURING:
PIRCON-PECK PROCESS**

by

R. B. Boyda
Arthur G. McKee & Company
10 S. Riverside Plaza
Chicago, Illinois 60606

FOREWARD

Information presented in this paper is intended to introduce the Pircon-Peck flue gas desulfurization process and highlight some of its attractive features. Because process developments are still generating patent applications, in addition to those already issued or still pending and allowed, many key details are not included. We invite anyone desiring additional information to contact Mr. Gene Barber in McKee's Chicago office at telephone number (312) 454-3899.

INTRODUCTION

The Pircon-Peck flue gas desulfurization (FGD) process presented in this paper was developed by Mr. Ladd Pircon and Dr. Ralph Peck. The process utilizes patented heterogeneous reactor technology developed by Mr. Pircon as cooling and absorber towers in conjunction with chemistry developed and demonstrated by Mr. Pircon and Dr. Peck at the Illinois Institute of Technology. The process is unique in that in addition to providing a means of controlling SO₂ emissions, it provides the owner with an opportunity to earn a profit on the invested capital.

PROCESS DESCRIPTION

The Pircon-Peck FGD process, as shown in Figure I, utilizes "activated" phosphate rock, ammonia, and flue gas as raw materials to produce ammoniated phosphate fertilizers. Laboratory and pilot plant testing indicate that in the process of producing fertilizer, sufficient SO₂ can be removed from the flue gas to provide compliance with all EPA regulations, both existing and proposed.

The chemistry utilized to produce the product material can be summarized by three chemical reactions. These are:

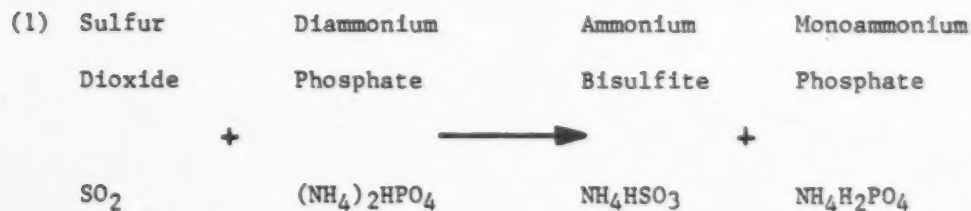
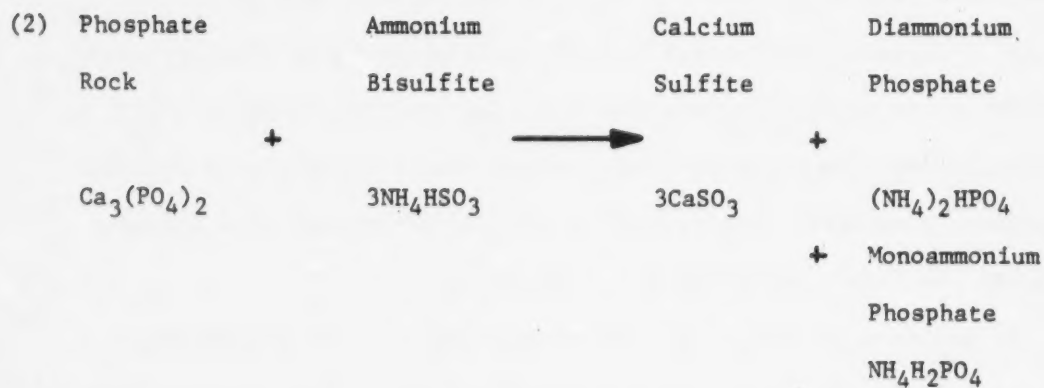
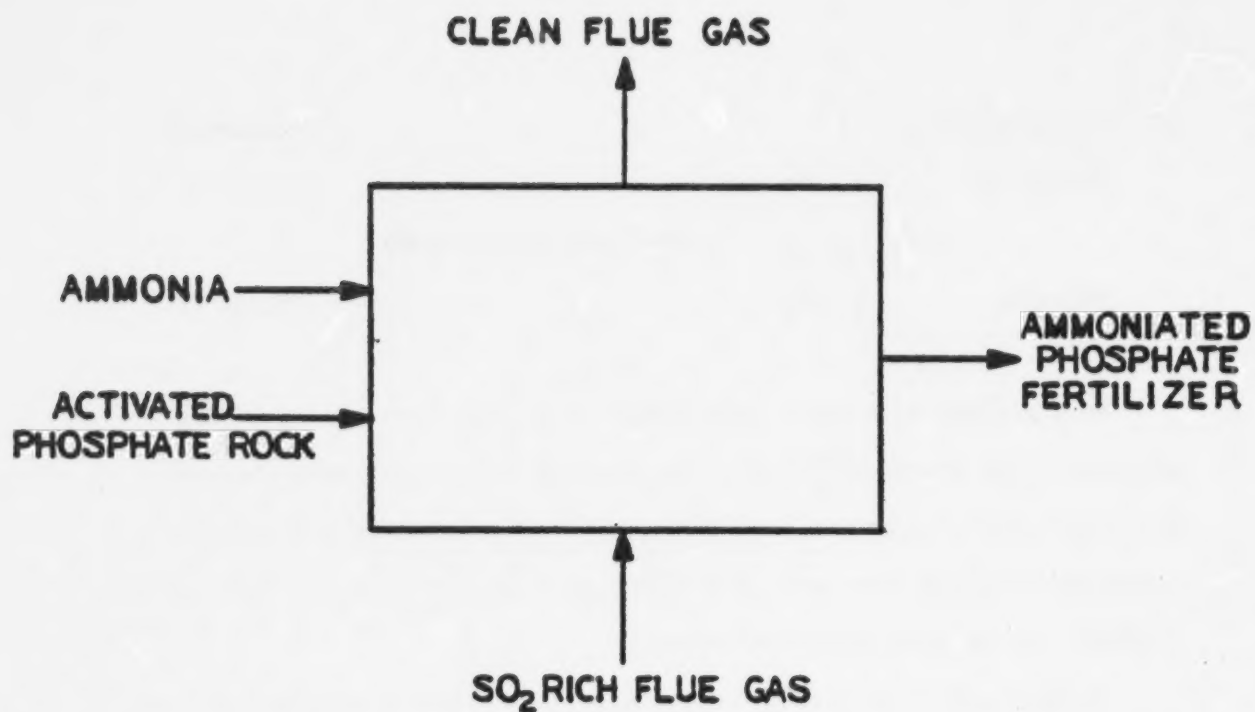
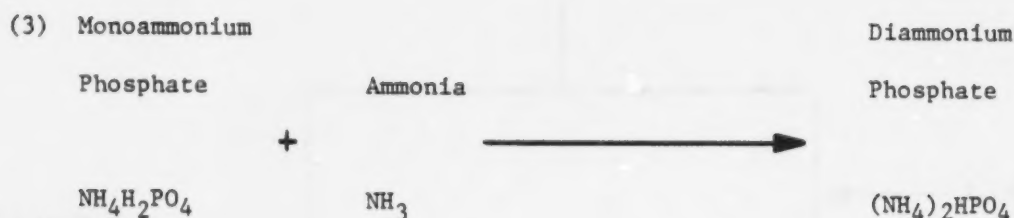


FIGURE I





An important feature of this chemistry is the pretreatment of the phosphate rock which facilitates the reaction of the rock with the weak sulfurous acid produced in the scrubber. Standard phosphate fertilizer chemistry requires concentrated sulfuric acid to acidulate the rock and liberate the desired phosphate molecule.

In many ways, the Pircon-Peck process is similar to standard double alkali technology. In each system, a soluble alkali is utilized in the scrubber portion of the process and in each system this alkali is regenerated by a calcium source. In the Pircon-Peck process, the soluble alkali is diammonium phosphate and the calcium source is phosphate rock. However, where alkali regeneration with lime produces water as a by-product, diammonium phosphate regeneration with phosphate rock produces additional phosphate. Neutralization of this by-product with ammonia produces the final diammonium phosphate product.

In addition to producing diammonium phosphate, the regeneration reaction also produces calcium sulfite and sulfate. This material can be included in the fertilizer product or separated for disposal. If included in the fertilizer product, the material would be similar to superphosphate fertilizers presently produced by the fertilizer industry.

These fertilizers contain the calcium sulfate solids produced by the acidulation of phosphate rock with sulfuric acid. While the final product form will be dictated by the local market conditions, the advantage of including the calcium sulfite/sulfate solids in the product is obvious.

Figure II provides a view of the scrubber portion of the process. Flue gas enters the system and is first passed through a high efficiency particulate control device. The use of a particulate removal system minimizes the possibility of product contamination from fly ash. After fly ash removal, the gas is sent to a cooling tower where it is quenched and saturated to its adiabatic saturation temperature. The cooled and saturated flue gas is next processed through the heterogeneous reactor tower (see Appendix I) where the flue gas is contacted with a saturated solution of diammonium phosphate and the SO_2 is removed. After passing through a demister, the SO_2 free flue gas is sent to the stack. If necessary, reheat can be added to aid in plume dispersion.

The SO_2 rich absorption liquor is taken from the heterogeneous reactor system and sent to an agitated attack tank where it is contacted with ammonia and phosphate rock. The slurry generated in the attack tank is next sent to a product separator (clarifier) where solid diammonium phosphate and calcium sulfite/sulfate crystals are withdrawn as a slurry. The product slurry is sent to a fertilizer plant where the product is converted to its final form. The overflow from the product separator is a saturated solution of diammonium phosphate and is sent back to the heterogeneous reactor tower for further reaction with SO_2 .

PIRCON-PECK FGD PROCESS

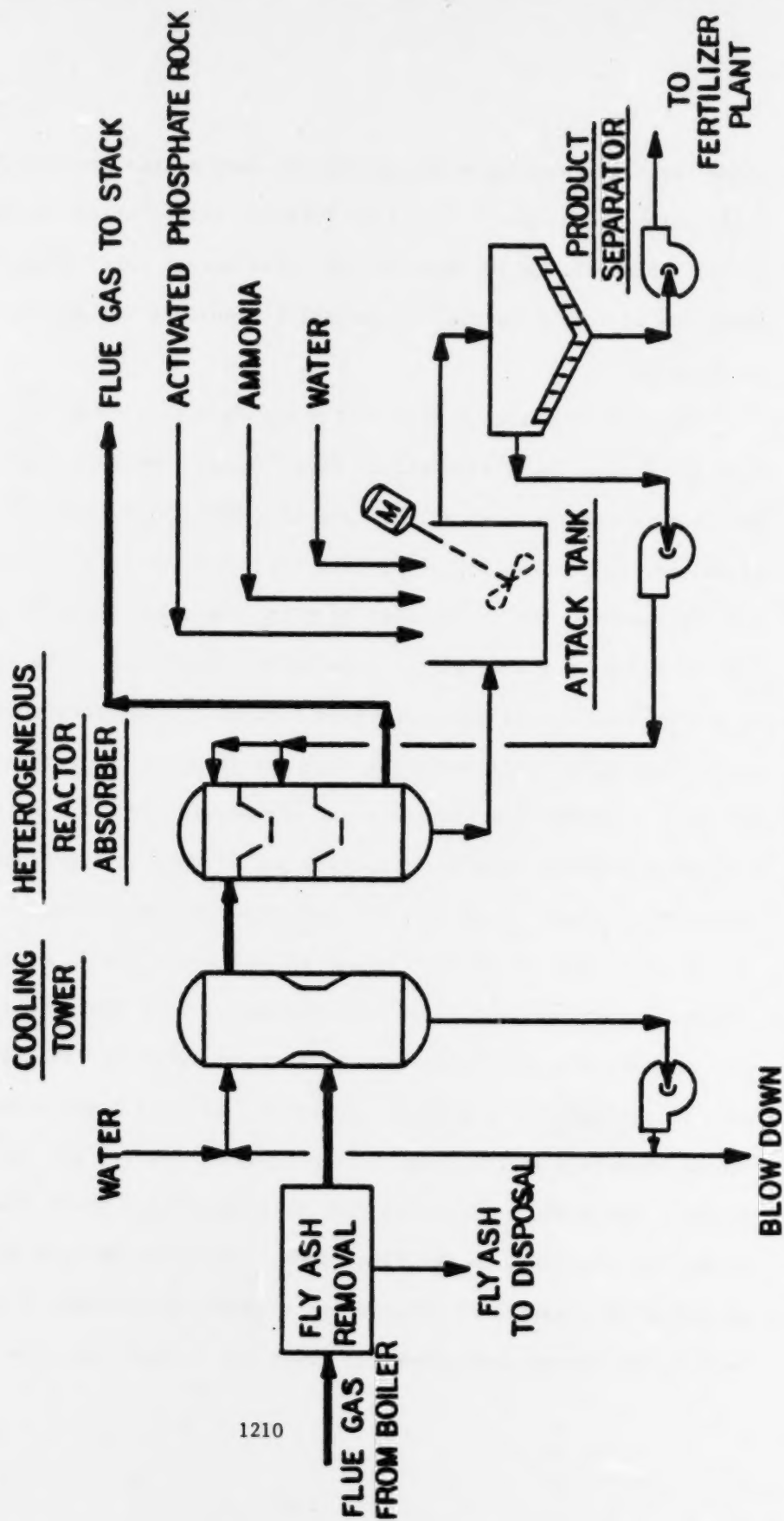


FIGURE II

The material from the product separator, on a dry basis, has an approximate analysis of 7-20-0 (N, P_2O_5 , K_2O). This analysis can be adjusted as desired by the addition of ammonia, phosphoric acid, or other N-P-K materials. The product form chosen by McKee for evaluation is a granular material with an analysis of 9-30-0. This analysis was chosen because posted prices are available to facilitate an economic evaluation of the process. Since fertilizer granulation plants are standard technology, no details are included.

As indicated, it is possible to produce product in various forms. Some alternates to granules with N and P_2O_5 values are:

1. Pure diammonium phosphate
2. Granules with N, P, and potassium values
3. Suspensions
4. Flakes

PROCESS HISTORY

The Pircon-Peck process evolved from Mr. Pircon's knowledge of phosphate chemistry, which he developed during his employment in the fertilizer industry, and his invention of the heterogeneous reactor technology. Mr. Pircon invented the heterogeneous reactor technology in the early 1970's and commercialized it as a low cost, high efficiency particulate control device. This technology was marketed commercially and over a dozen installations are presently in operation.

The major event in the development of the technology occurred when funding was obtained from the Illinois Institute for Environmental Quality for the design, construction and operation of an 800 acfm pilot plant. This pilot plant was installed at the Illinois Institute of Technology (IIT) and operated by graduate students for more than two years. This work was supervised by Dr. Ralph Peck. Work performed by students resulted in IIT granting two doctorate and six masters degrees in chemical engineering.

The operation of this pilot plant provided confirmation of the process principles and information for design of a commercial sized demonstration plant. In addition, sufficient fertilizer product was produced so that test work could be performed by Argonne National Laboratories. This test work was also funded by the Illinois Institute for Environmental Quality.

At the conclusion of the pilot plant work, the following items were demonstrated:

1. More than 95% of the sulfur oxides present in flue gases generated by a 6.2% sulfur coal, could be removed by the pilot plant scrubbing system.
2. Particulates generated by the underfeed stoker fired boiler were removed so that no particulates were visible in the stack plume.
3. The process produced a satisfactory fertilizer product.
4. The required liquid/gas ratios for the system were considerably less than those for conventional scrubbers.
5. The overall pressure drop across the pilot plant system, including the boiler, cooling tower and scrubber, was about two inches of water.

PROCESS EVALUATION

McKee's evaluation of the Pircon-Peck process was performed by reviewing laboratory data, pilot plant records and pilot plant operating experience. In addition, extensive discussions were held with the process inventors to develop additional background information. This investigation has shown that:

1. The process can be carried out in a plant of very simple design.
2. Few moving parts are involved in the equipment.
3. There is considerable flexibility in the process.
4. The process can be run with clear liquids or slurries in the heterogeneous reactor tower.
5. The product fertilizer analysis can be adjusted to fit the needs of the local market.
6. Scaling in the absorber is not a problem due to the process chemistry.
7. For the same removal rates, equipment is simpler and smaller than for other scrubbing processes.

The economics of the process were evaluated by preparing a mechanical design and a heat and material balance for the 100 mw sized facility. Once this was complete, capital and operating cost estimates were made and an economic analysis performed. The design basis used for this work was:

Fuel: 3.5% Illinois High Volatile C Coal @ 11,500 Btu/lb.
Heat Rate: 11,500 Btu/kwh
Flue Gas: 2.12 scfm/kw @ 75° F
SO_x Removal: 95%
Phosphate Rock Conversion: 95%
Service Factor: 85% (7450 Hr./Yr.)

On this basis, it was estimated that the plant, including the scrubber and fertilizer granulation plant, would cost approximately \$10.7 million and produce 128,600 tons per year of a 9-30-0 analysis granular fertilizer product. It was estimated that the margin between the fertilizer manufacturing cost and selling price would be \$50.00 per ton.

Using these economic parameters, a project feasibility analysis was performed. This analysis was based on:

- . Discounted Cash Flow Techniques
- . Twenty Year Project Life
- . Income Tax @ 50%
- . Sum of the Years Digits Depreciation
- . Working Capital of 13% of Sales

This study indicated that the project would have an Investors Rate of Return (ROI) of 22% over the life of the project.

Because of the preliminary nature of the estimate and the inherent uncertainty of prices in the fertilizer industry, McKee believes that the significance of the indicated ROI is not its absolute value, but the fact that it is significantly positive for even a small sized facility.

The economic viability of the process was further confirmed by studies of the sensitivity of ROI to changes in key economic parameters. Parameters studied included capital cost, production volume, variable costs and margin. As an example of this work, Figure III shows the sensitivity of ROI to variations in the margin between selling price and manufacturing cost. These analyses show that this margin is the most sensitive parameter. A summary of the projected operating costs is given in Figure IV.

McKee has also looked at the market for the process product. At the present time the market for fertilizer, on a P_2O_5 basis, exceeds 10 million tons per year and is expanding at a rate of approximately 6% per year. On the design basis presented in this paper, a 500 mw plant would produce 190,000 tons per year of P_2O_5 equivalent product which would fit easily into the growing fertilizer market. This is particularly true in the midwest and farm belt regions.

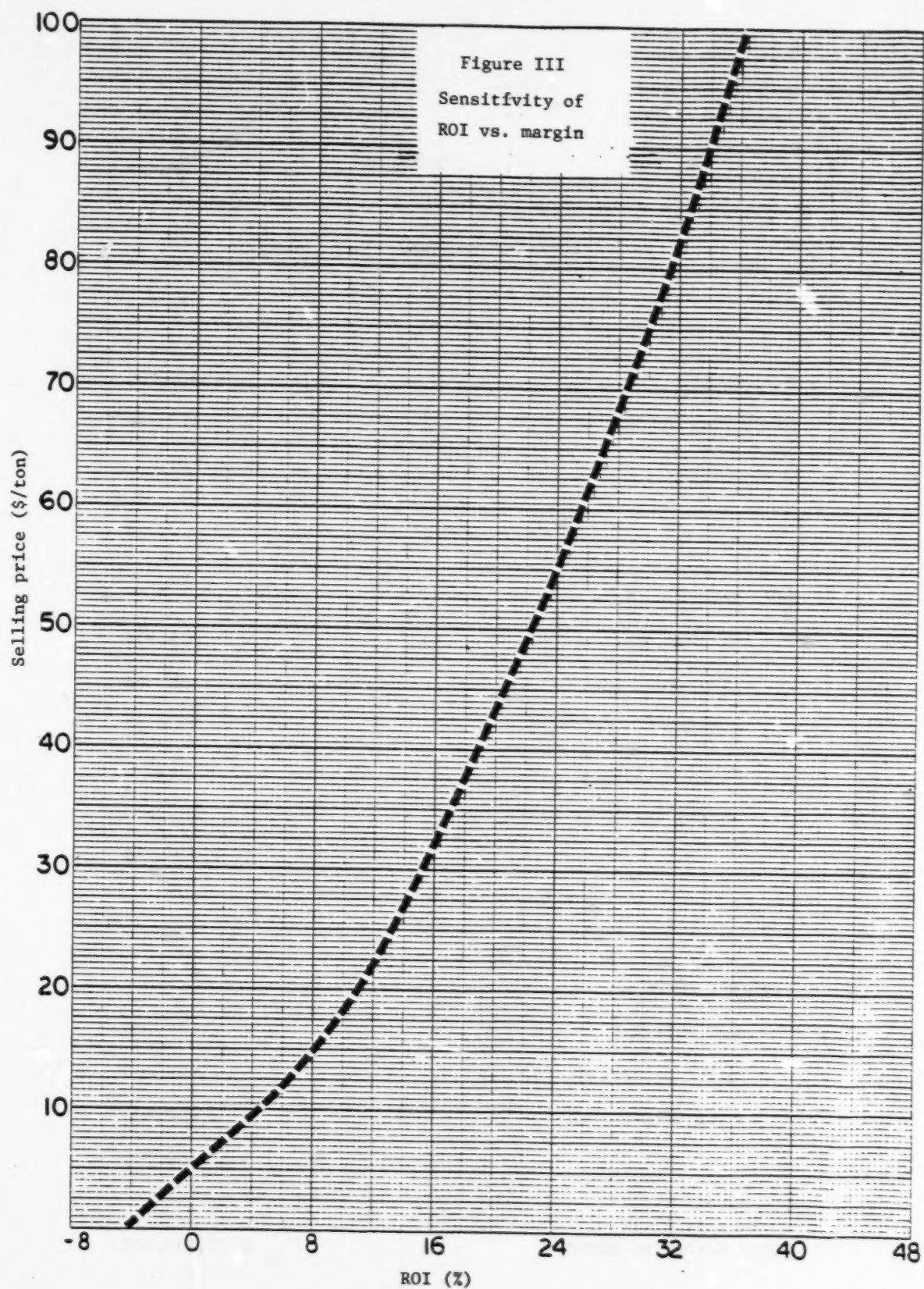


FIGURE IV

CAPITAL & OPERATING COST SUMMARY

INVESTMENT	\$ 10,726,000
PRODUCTION VOLUME	128,635 Tons/Yr.
SALES MARGIN	\$ 50 Ton
SEMI VARIABLE COSTS	
MAINTENANCE	\$ 845,000
LABORATORY	96,000
INSURANCE & PROPERTY TAX	370,000
PLANT OVERHEADS	534,000
ADMINISTRATION & MARKETING	<u>294,000</u>
	\$ 2,139,000
VARIABLE COSTS	
RAW MATERIAL	\$ 8,225,000
LABOR & SERVICES	1,009,000
UTILITIES & FUEL	<u>924,000</u>
	\$ 10,158,000
TOTAL OPERATING COSTS	\$ 12,297,000

CONCLUSION

As a result of our process studies McKee believes that the Pircon-Peck process offers significant advantages over existing commercial processes. Not only does the process provide an opportunity to make pollution control profitable, it actually provides an economic incentive to utilize high sulfur coals. This incentive results from the fact that the quantity of fertilizer product produced is directly proportional to the quantity of sulfur captured. As a result, the plant operator will want to operate the system at SO₂ capture rates in excess of EPA standards.

McKee has recently entered into an agreement with the process owner, Mr. Ladd Pircon, to commercialize the technology. It is our present co-objective to accomplish this by the design, construction, and operation of a commercial sized demonstration plant with a minimum rating of 100 mw. At the present time we are working to put together the four items necessary to achieve our objective. These four items are; funding, a site for the project, an operator with fertilizer production experience and a commitment from a fertilizer company to market the fertilizer product. Significant progress has been made in this effort at two separate locations.

APPENDIX I
HETEROGENEOUS REACTOR
TOWER

The heterogeneous reactor tower included in this design utilizes differential velocities between liquid, gas, and solid particles to obtain high mass transfer coefficients. The tower internals (see Figure V) consist of a series of two conical restrictions preceded by spray nozzles and followed by impinger plates. Each plate is washed by a dedicated spray nozzle.

The differential velocities are achieved at a low absolute velocity as the gas enters the conical restrictions. Because the gas phase is compressible it accelerates more rapidly in the cones than the solid or liquid particles which are accelerated by drag forces from the gas. When the gas exits the conical restriction and impinges on the plate, the streamlines of the gas cross those of the droplets and the solid particles and again produce a high contact efficiency.

As indicated in the process history section this technology has been commercialized as a low cost high efficiency control device. This success with large scale units, plus field tests of the process chemistry, provides strong indications that the demonstrated success of the pilot plant scrubber can be duplicated in commercial units.

HETEROGENEOUS REACTOR

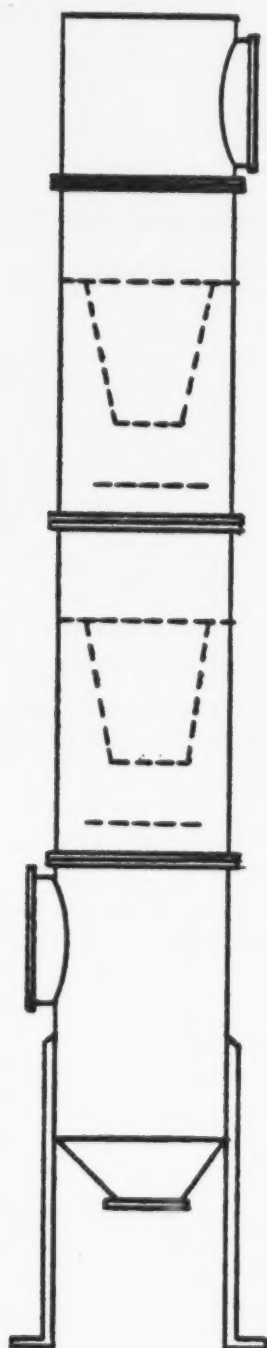


FIGURE V

**DRY FGD AND PARTICULATE CONTROL
SYSTEMS**

By

K. A. Moore

R. D. Oldenkamp

**Rockwell International
Energy Systems Group**

M. P. Schreyer

Wheelabrator-Frye, Inc.

D. W. Belcher

Stork Bowen Engineering, Inc.

DRY FGD AND PARTICULATE CONTROL SYSTEMS

INTRODUCTION

For the past 8 years, Rockwell International has been developing advanced systems for flue gas desulfurization. The regenerative Aqueous Carbonate Process for Flue Gas Desulfurization* uses a spray dryer as a flue gas contactor and generates dry reaction products which are collected in a particulate removal device, regenerated chemically and reused. A simplification of the process, wherein the regeneration system is not used, has also found acceptance as the Two Stage "dry scrubbing" Process.** This process was developed jointly by Rockwell and Wheelabrator-Frye and tests have shown the dry, two-stage system's capability to remove both SO_2 and particulates to levels which meet existing and proposed environmental standards. Recent tests have shown that a variety of alkali compounds can be used in solution or slurry form in a spray dryer contactor to remove SO_2 from boiler flue gases. The dry particulates that leave the spray dryer, such as fly ash, alkali sulfates, sulfites, and unreacted alkali, are removed from the flue gas by a fabric filter and disposed of as a dry powder.

Tests have shown the dry, two-stage system's capability to remove both SO_2 and particulates to levels which meet existing and proposed environmental standards.

This past year, three "dry scrubbing" systems, totaling 1450 MW of capacity, have been contracted for. Each system features different alkali injection and product collection concepts to achieve dry scrubbing. This paper proposes to review the basic concepts and design features of the three different systems. Major emphasis will be given to the Rockwell/Wheelabrator Two Stage Process.

*U.S. Patent 3,932,587

**Patent Pending

DRY FGD PROCESS DESCRIPTION

The process of SO_2 removal utilizing spray dryers is inherently simple; see Figure 1. The process features the wet contact of SO_2 in flue gas with a fine mist of water containing a solution or slurry of alkali, typically either lime or soda ash. The SO_2 is absorbed in the mist water droplet and neutralized by the alkali. With boiler flue gas nominal temperatures of 300°F , the small quantity of water (about 0.3 gal per 1000 acf) in the alkali mist is completely evaporated. Thus, the flue gas is not saturated and remains warm. The exit gas temperature from the spray dryer is 150° to 200°F , a temperature safely above the water dew point. This technique results in particles of dry, spent alkali which have captured the SO_2 . Essentially all of the particulate, including the fly ash, is then removed in product collection equipment (fabric filter or electrostatic precipitator). SO_2 removal efficiencies of 85% or more are achievable with particle emissions to the stack of less than 0.03 #/mBtu (~ 0.01 grain/scfS).

There are many benefits to keeping the flue gas warm and dry. The materials of construction can be of carbon steel with minimum risk of erosion, corrosion, or scaling. Spray dryer designs prevent wetting the walls of the vessel, and the dry alkali powder provides a renewable protective dust coating on chamber and duct surfaces. The flue gas is warm and dry at the exit of the product collection device; thus, reheat is not normally required to achieve plume dispersal out of the stack. The ID fan, downstream of the product collector, operates in a clean, low-temperature environment.

Water chemistry and PH controls typical of wet scrubbing systems are eliminated. The spray dryer exit temperature is the preferred primary signal for control of total water feed to the system. The alkali feed concentration is determined by the SO_2 emission requirements.

Additionally, the resulting dry product eliminates the disposal problem of sludge ponds. The product is a free-flowing material that can be returned to the coal mine or another pit for final disposal.

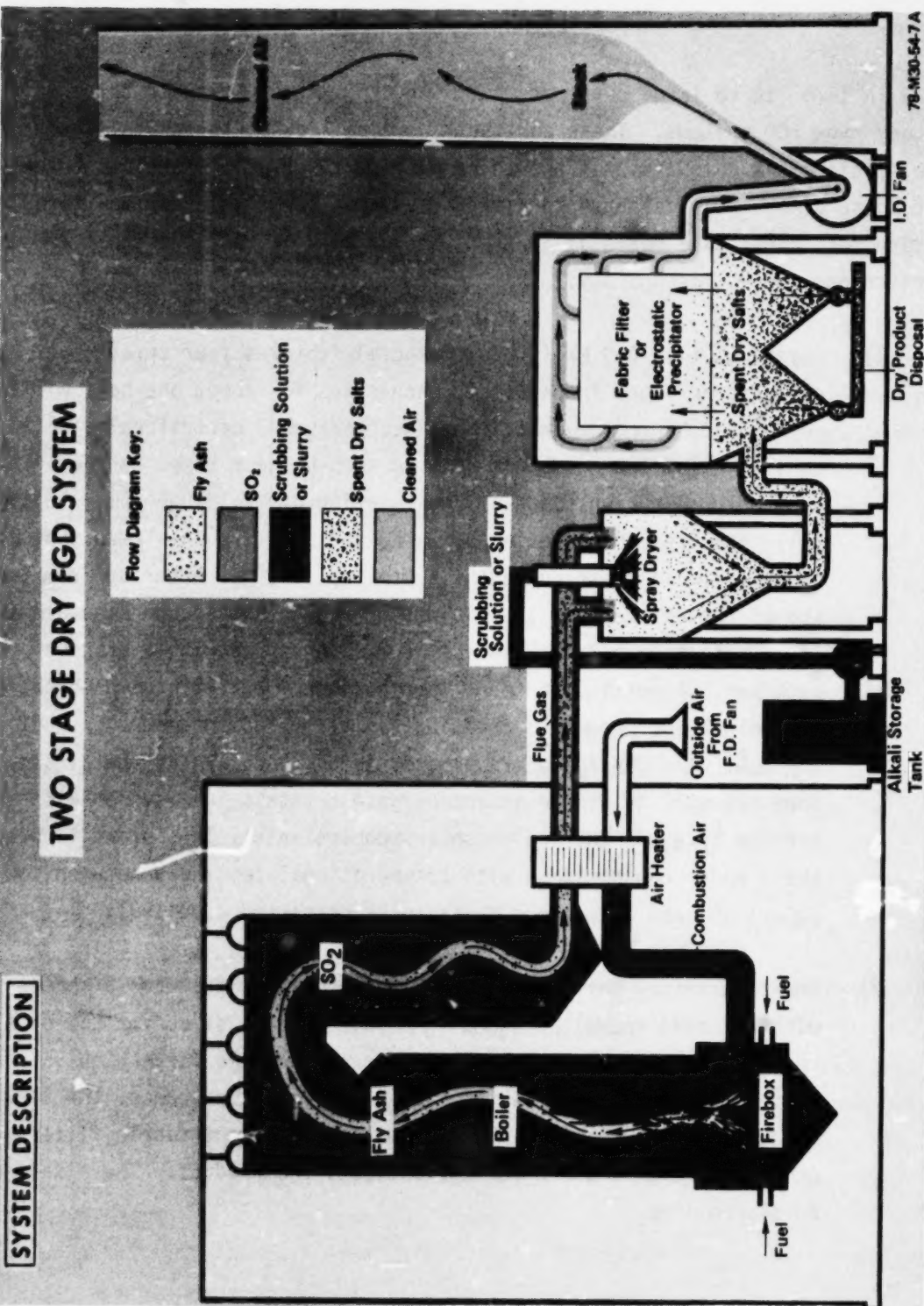


Figure 1. Two Stage Dry FGD System

DRY SCRUBBING SYSTEMS SUMMARY

In 1978, three large contracts totaling 1450 MW were awarded to suppliers of dry-type FGD systems. These contracts were all for North Dakota lignite coal applications. A comparison of the three different scrubbing concepts selected for these plants provides an interesting review of the development of spray drying technology as a means of SO₂ removal and also illustrates the trends in particulate control technology.

- 1) Coyote Unit 1, 410 MW (1,890,000 acfm) features four spray dryer chambers arranged in parallel, each sized for about one-half million acfm. Each spray dryer chamber incorporates 3 centrifugal atomizers, consisting of 150 hp motors driving step-up gear boxes turning a disk or wheel. The high speed of the rotating disk, 18,000 rpm, centrifugally atomizes the solution of water and alkali. This resulting mist is injected in cross-current flow to the flue gas entering from the top of the chamber and distributed by vane rings around each atomizer (see Photo Figure 2.) Each chamber is furnished with a standby atomizer. A multicompartment fabric filter, utilizing a combination mechanical and pneumatic cleaning cycle, collects the dry product and fly ash. The Coyote owners elected to use soda ash as the alkali. Soda ash will be stored as monohydrate crystals in a system with a minimum of equipment and manpower requirements. This plant is presently under construction with preoperational testing scheduled for July 1980, and acceptance tests to be completed June 1981.
- 2) Antelope Valley Unit 1, 440 MW (2,200,000 acfm) features 5 chambers (4 with a ducted spare) arranged in parallel, each sized for 25% of the volume. In this design, the spray dryers use one large high-horsepower rotary atomizer per chamber. For maintenance, the flow is switched to the spare chamber. A multicompartment fabric filter, utilizing reverse air cleaning, collects the product. The alkali will be slaked lime.

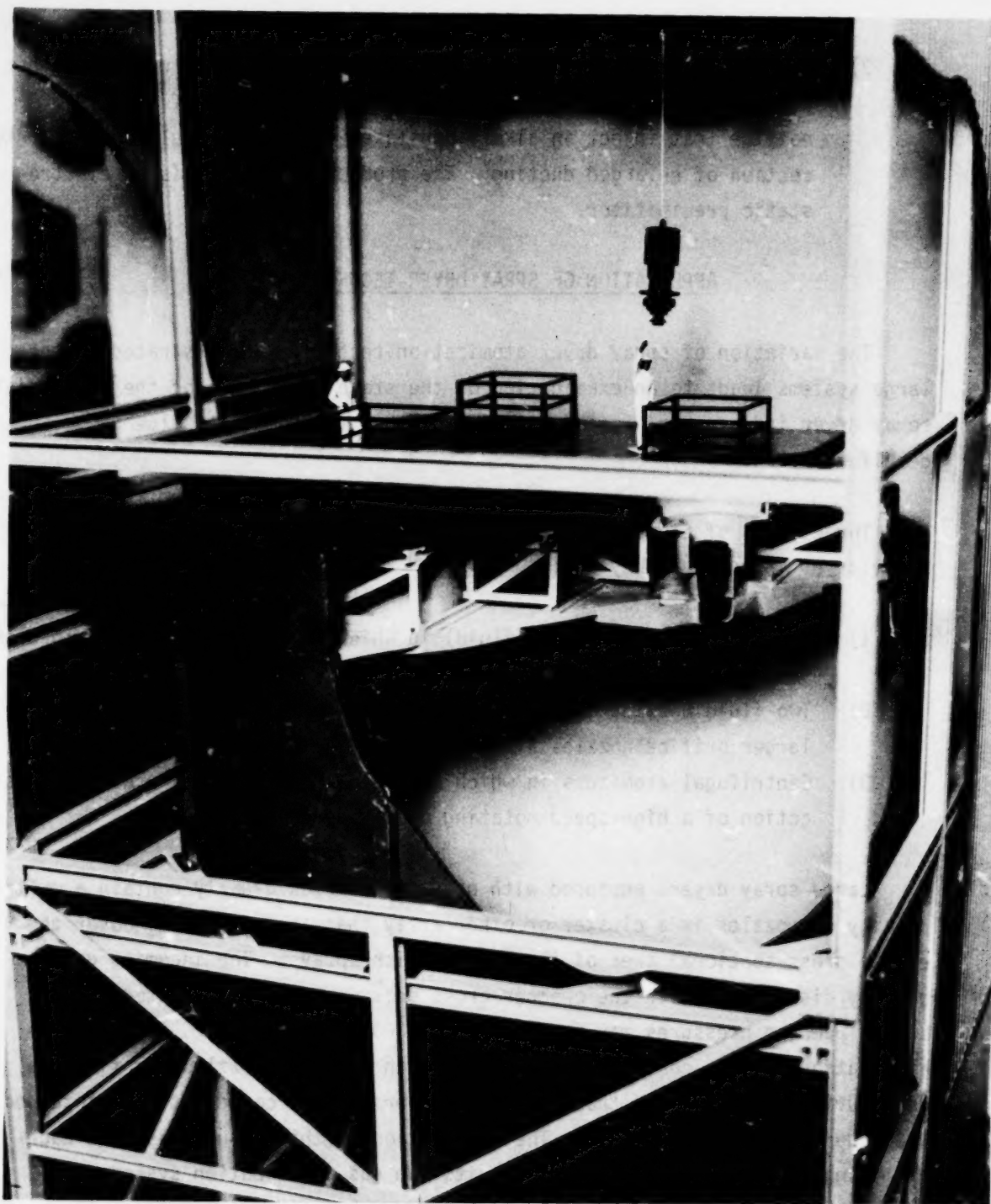


Figure 2. Spray Dryer With Three Atomizers

- 3) Laramie Station Unit 3, 600 MW (2,810,000 acfm) features yet another design. This "spray dryer" utilizes a large number of two-fluid nozzles that inject an alkali of slaked lime into a multicompartmented section of enlarged ducting. The product collection is by electrostatic precipitator.

APPLICATION OF SPRAY DRYER TECHNOLOGY TO FGD

The variation of spray dryer atomization technology demonstrated by these large systems leads to an examination of the state of the art of the commercial spray dryer industry and the application of this technology to flue gas desulfurization.*

Three types of atomizers have been used successfully in the commercial spray dryer industry. They are:

- 1) Pressure nozzles (single fluid) in which the feed is atomized through small orifice nozzles with a high-pressure pump.
- 2) Two-fluid nozzles in which the feed is atomized through relatively larger orifice nozzles by the action of a second pressurized fluid.
- 3) Centrifugal atomizers in which the feed is atomized by the mechanical action of a high-speed rotating atomizing device.

Large spray dryers equipped with pressure nozzles usually contain a multiplicity of nozzles in a cluster or other array that is designed to cover the entire cross-sectional area of the chamber with sprays. The incoming gas is usually distributed over the chamber cross section by one perforated plate. Liquid pumping pressures may be well above 1000 psi in order to achieve fine atomization. Tests conducted by Rockwell with pressure nozzles for flue gas desulfurization indicated that SO_2 removal performance could not be maintained when the gas flow was reduced. The implication of this result was that each atomizing device should be mated with its own gas distribution system if acceptable performance was to be maintained over a wide range of gas flows.

* The design features of the Coyote FGD System of Rockwell/Wheelabrator are discussed below.

Since pressure nozzles typically have a small capacity and limited turndown, the gas distribution problem could become quite complex. Also, pressure nozzles are not suitable for abrasive slurry service. Thus, this type of atomization was rejected by Rockwell for the general FGD application.

Large spray dryers equipped with two-fluid nozzles are designed in the same fashion as described above for pressure nozzles. However, two-fluid nozzles have inherently greater capacity than pressure nozzles, and the relatively larger orifice is somewhat less susceptible to abrasion. The energy required for equivalent atomization is typically 50 to 100% greater than that for pressure nozzle or centrifugal atomization. Tests conducted by Rockwell with two-fluid nozzles for the FGD application indicated that it could be an operable technique if gas distribution could be optimized and maintained over the expected range of boiler loads, and if performance deterioration from nozzle erosion could be minimized. Even with solutions to these problems, the energy penalty tends to make two-fluid nozzles less desirable than the centrifugal concepts for the general FGD application.

Large spray dryers equipped with centrifugal atomizers have been used in the spray dryer industry for most slurry services. A single, high-capacity atomizer is matched to a "vane ring" gas distribution device. The trend has been to develop increased capacity for single atomizers rather than to build dryers with multiple atomizers. This trend is specific to the normal spray drying applications where gas flow is minimized and feed rate is maximized. Also, many applications require specific product characteristics that could not be assured if multiple atomizers were employed. There is no technical reason that multiple centrifugal atomizers cannot be used in a single large chamber for the FGD application. In fact, Masters (of Niro Atomizer) in his book Spray Drying states that, "Although one atomizer unit is used per drying chamber, there is no reason why multi-atomizer units cannot be applied in very large drying chambers. Where multi-atomizer units are used, drying air is supplied around each atomizer wheel."

One of the largest spray drying chambers ever built (~ 50-ft diameter) contained multi-atomizers of the centrifugal type. It was built in 1930 by Bowen Engineering for a potato drying application where product particle size was not a key criteria. It operated successfully. More recently, Bowen (now Stork-Bowen) and Rockwell have tested three centrifugal atomizers in a single, 7-ft-diameter pilot spray dryer. This provided a severe test of the concept since maximum spray pattern overlap was to be expected in the small chamber. The essential experimental results were:

- 1) For fixed L/G ratios, three atomizers with three vane rings gave equivalent SO_2 removal (~90%) as that observed when a single, larger atomizer was tested in the same chamber.
- 2) When one of the three atomizers was shut off and the L/G was preserved by increasing fluid flow to the two remaining in operation, acceptable SO_2 removal was maintained (~89%).
- 3) Even when two of the three atomizers were shut off (again preserving the L/G), the SO_2 removal dropped to about 82%. In all cases, the gas continued to flow through the three separate vane rings.

It would be expected that SO_2 removal performance would be affected somewhat more if three larger atomizers were tested in the same manner in a large chamber. However, if the gas were diverted from a nonoperational atomizer to those which were in operation, no serious performance deterioration should be observed. In fact, agglomeration of the droplets will result in larger product particle size which will be a positive factor in its collectability.

The key for FGD applications is to provide adequate gas distribution to each of the operating atomizers. If this is accomplished, there should be no concern about spray pattern overlap. Of course, adequate residence time must be provided to ensure that the agglomerated droplets are fully dried.

Recent flow model tests (1/16 scale) of the Coyote FGD system have provided excellent indications of the operability of the gas distribution system designed for four 46-ft-diameter chambers with three atomizers per chamber. Under all

gas flow conditions from 25 to 100% of design flow, the gas distributed equally to each of the four chambers. Specifically, the individual chamber flows were $25 \pm 2\%$ of the total system flow over the entire flow range. The distribution of gas to each of the three atomizer vane rings in the four chambers was found to be $33 \pm 2\%$ of the flow to that chamber. This again was tested over the full flow range. The results of this testing should not be surprising, since it is relatively easy to divide the rectangular gas inlet duct of a Stork-Bowen spray dryer into three equal cross-sectional areas.

It should also be noted that, although standard spray dryers have been built with centrifugal atomizers of 400 hp and above, no dryers are currently in service which handle gas volumes of greater than about 200,000 acfm (at dryer outlet conditions). Thus, the use of a single centrifugal atomizer in an FGD application with 400,000 acfm (again dryer outlet conditions) or greater gas flows constitutes a significant departure from proven gas distribution (atomizer-vane ring) spray dryer technology. Conversely, the use of three centrifugal atomizers, each handling 1/3 of the gas flow, is within the experience level of atomizer-vane ring design.

Considering all of the above factors, tests, concepts, and experience, the Rockwell-Wheelabrator Joint Venture and Stork-Bowen have designed the Coyote FGD system with the following features:

- 1) Each of the four 46-ft chambers will contain three 150-hp atomizers with matching vane ring gas distributors. In the event of an atomizer malfunction, the feed can be diverted to the two operating atomizers to sustain SO_2 removal performance. The atomizer motors are specifically oversized to allow this option.
- 2) All four spray dryer chambers are connected in parallel for total gas flow. Separate ducts and dampers are provided to each atomizer. Therefore, gas flow control is available to assure high SO_2 removal levels for the range of boiler loads.

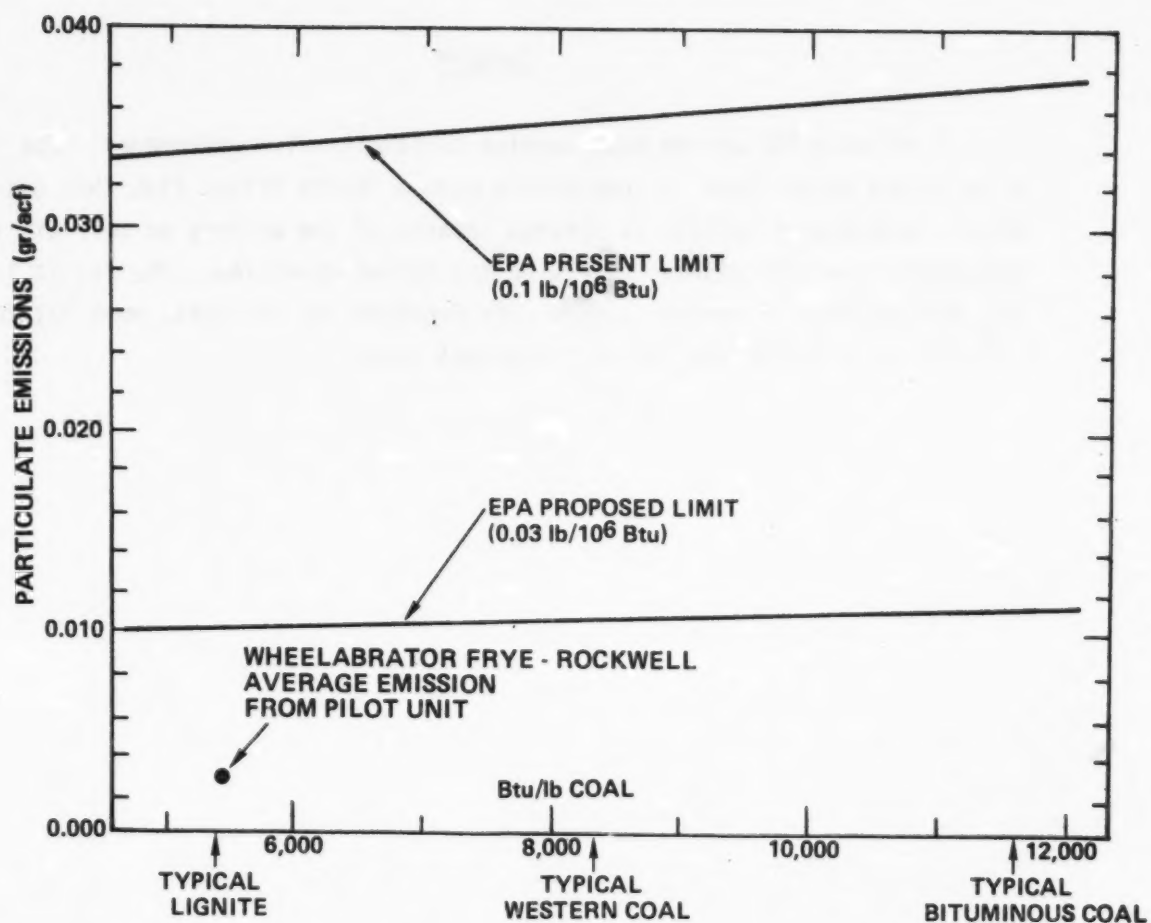
- 3) A spare atomizer is provided for each chamber to permit rapid replacement in the event of malfunction or atomizer removal for maintenance purposes. Therefore, no spare chamber is required to ensure that the FGD system is fully available for service.
- 4) The chambers have been sized with at least a 50% greater residence time than that proven by 9 years of pilot test experience. This will protect against inadequate drying because of possible agglomeration.
- 5) Atomizer position in the large chambers was specifically designed to preclude impingement of wet droplets on the chamber walls. This prevents chamber wall buildup.

PARTICULATE CONTROL

The proposed new source performance standards for particulates are being significantly tightened. Figure 3 reveals the impact the requirement will have on equipment suppliers who must design the particulate control equipment to exceed typical NSPS requirements by a comfortable margin. For many fuels, high resistivity fly ash precludes the economic use of an electrostatic precipitator. For the Coyote plant, a multicompartment fabric filter will provide reliable particulate control. Figure 3 indicates the results of typical fabric filter performance during pilot tests with lignite for the Coyote program.

As fabric filters for power boilers are quite large, a number of factors need to be considered to achieve economical design and operation. The effect of the spray dryer upstream of the fabric filter is to reduce the gas volume (reduced temperature) but increase the dust load.

The increased dust load requires a positive approach to bag cleaning. For Coyote, the fabric filter will use a combination mechanical shake and reverse air cleaning cycle to remove the heavy filter cake and operate at a minimum system pressure drop. In pilot tests conducted by WFI, the combination cleaning proved superior to reverse air only.



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Figure 3. Particulate Emission Control Requirements

The lowered flue gas temperature provided by the upstream spray dryer allows the fabric filter to be smaller and to utilize a long-lived acrylic or polyester bag fabric instead of fiberglass. These features of fabric filter design result in fewer filter compartments, and fewer, less expensive bags.

SUMMARY

A reliable FGD system must combine simplicity with redundancy. The use of a multihead spray dryer in combination with a fabric filter fits this description. System availability is insured because of the ability to maintain the equipment "on-line" without affecting the boiler operation. The result is an SO₂ and particulate control system that provides the simplest, most reliable solution to emission control on the market today.

ATTENDEES

Flue Gas Desulfurization Symposium
Las Vegas, Nevada
March 5-8, 1979

Abdulsattar	Abdul Hameed	50 Beale Street	San Francisco	CA	94119	Bechtel National Inc.
Abrams	Jack	50 Beale Street	San Francisco	CA	94105	Bechtel National, Inc.
Achenbach	D.	One Penn Plaza	New York	NY	10001	Chemico Air Poll. Control
Achtner	S.	One Penn Plaza	New York	NY	10001	Chemico Air Poll. Control
Adams	R. L.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye Inc., APCD
Adams	Radford C.	201 W. Roxboro Street	Durham	NC	27701	TRW Inc.
Adams	Robert A.	77 Havemeyer Lane	Stamford	CT	06904	Dorr-Oliver, Inc.
Ager	James L.	1200 6th Avenue	Seattle	WA	98101	W. States Reg. Coordinator, 208 Program
Ahman	Stefan	S-35187 VASJO	Seattle	WA	98101	Flakt Industrial Division
Aikawa	H.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Albrandt	Alan	5900 E. 37th Avenue	Denver	CO	80207	Public Service Company
Alfredsson	Per-Olo	AB Svenska Flaktfabriken	Stockholm	SWEDEN		General Electric Company
Alger	John R. H.	Building 2-447	Schenectady	NY	12345	Public Service Indiana
Ali	Sy	1000 E. Main Street	Plainfield	IN	46168	Evans L. Shuff & Associates
Alaand	Charles F.	6303 Barfield Road, Ste. 219	Altanta	GA	30328	Chuo University
Ando	Junpei	1-13 Kasuga, Bunkyo-Ku	Tokyo	JAPAN	112	Babcock & Wilcox
Andresen	R. L.	1545 Wilshire	Los Angeles	CA	90017	Industrial Clean Air
Andrew	Rob	300 Civic Drive	Pleasant Hill	CA	94524	Pacific Power & Light Company
Angelovich	John M.	920 SW 6th Avenue	Portland	OR	97204	Gilbert/Commonwealth
Ansari	Armad H.	209 E. Washington Avenue	Jackson	MI	49201	McIlvaine Company
Ardell	Marilyn	2970 Maria Avenue	Northbrook	IL	60062	Lurgi Umwelt und Chemotechnik GmbH
Aras	Karlheinz	Gervinusstrasse 17/19 6000	Frankfurt/Main	GERMANY		Lodge Cottrell Ltd.
Ashley	Michael J.	George Street Parade	Birmingham, B310Q	ENGLAND		U.S. Steel Corporation
Auger	Robert	125 Jamison Lane	Monroeville	PA	15146	PEDCO Environmental Inc.
Avinash	F.	11499 Chester Road	Cincinnati	OH	45246	Research Triangle Institute
Ayer	Franklin A.	P. O. Box 12194	Res. Tri. Park	NC	27709	Lower Colorado River Authority
Babcock	John E.	P. O. Box 220	Austin	TX	78767	Omaha Public Power Distri.
Bachman	Gerald G.	1623 Harney Street	Omaha	NE	68104	IU Conversion Systems, Inc.
Bacskai	Ronald J.	115 Gibraltar Road	Horsham	PA	19044	Industrial Clean Air
Badger	Bill	300 Civic Drive	Pleasant Hill	CA	94524	Barber-Greene Company
Baer	Richard E.	600 N. Highland Avenue	Aurora	IL	60507	A-S-H PUMP Div. Envirotech Corp.
Bailey	Albert E.	33 Sproul Road	Frazer	PA	19355	Apollo Chemical Corp.
Bailey	Ann	35 South Jefferson Road	Whippany	NJ	07981	ARCO Petroleum Products Co.
Bains	C. S.	400 East Sibley Blvd.	Harvey	IL	60426	Envirotech Corp.
Balakrishnan	N. S.	4233 N. United Parkway	Schiller Park	IL	60176	Envirotech Corp.
Ball	Hyron L.	400 Rouser Road	Pittsburgh	PA	15108	Federal Gov't.-Dept. of Environment
Bambrough	H. A.	351 Boulevard	St. Joseph, Mull	Quebec	CANADA	King Wilkenson Inc.
Banchik	N.	3701 Kirby Drive	Houston	TX	77098	Arthur G. McKee & Company
Barber	G. D.	10 South Riverside Plaza	Chicago	IL	60606	U.S. EPA
Barber	Walter C.	Mail Drop-10	Res. Tri. Park	NC	27711	TOSCO
Barnes	Bruce	Avon Refinery	Martinez	CA	94553	The CADRE Corporation
Bartlett	Terry W.	10 UOP Plaza	Atlanta	GA	30247	UOP - Air Correction Division
Basciani	Kenneth A.	200 S. Michigan	Des Plaines	IL	60016	Pullman Power Products
Bass, Jr.	Loren O.	20 Turnpike Road	Chicago	IL	60604	New England Power Service Co.
Batra	Sushil K.	MD-12	Westborough	MA	01581	U.S. EPA
Bauman	Robert D.	MD-12	Res. Tri. Park	NC	27711	Northrop Services, Inc./APTI
Beachler	David S.	MD-12	Res. Tri. Park	NC	27709	Wallace & Tiernan Div./Pennsalt Corp.
Beals	Joseph L.	2540 S. 27th Avenue	Broadview	IL	60153	Foster Wheeler Energy Corporation
Beckman	Eugene B.	180 Knoll Road	Roonton	NJ	07005	Carborundum Company
Beeler	Linda	P. O. Box 87	Knoxville	TN	37901	

Behie	S. W.	900-One Palliser Square	Calgary, Alberta	CANADA	T2G0R6	Montreal Engineering Company, Ltd.
Belcher	Donald W.	P. O. Box 598	Somerville	NJ	08876	Stork Bowen Engineering Inc.
Berst	A. W.	P. O. Box 2206	Birmingham	AL	35201	Zurn Industries
Bhatti	Mohamed R.	1300 Greenway Plaza E.	Houston	TX	77046	Pullman Kellogg
Biedell	Edward L.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell
Bielawski	G. T.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Company
Bierbower	R. G.	P. O. Box 1130R	Horseshoe	NJ	07960	Allied Chemical
Binkley	M. J.	P. O. Box 226226	Dallas	TX	75266	Glitsch, Inc.
Biondo	Samuel J.	1100 Independence Avenue	Washington	DC	20545	Department of Energy
Birkner	V. B.	P. O. Drawer 5000	Lakeland	FL	33803	Davy Powergas, Inc.
Bitsko	Ralph	2301 Market Street	Philadelphia	PA	19101	Philadelphia Electric Company
Blair	Gary T.	P. O. Box 932	Elkhart	IN	46515	Miles Laboratories, Inc.
Blau	Ted	Black Horse Lane	North Brunswick	NJ	08902	Ted Blau Associates
Blinckmann	R.	One Penn Plaza	New York	NY	10001	Chemico Air Poll. Control
Bloss	R. Edward	700 N. 6th Street	Levanon	PA	17042	Envirotech
Blythe	William B.	576 Azalea Road, Ste 124	Mobile	AL	36609	Envirotech Corporation
Bond	George	420 Rouser	Corapolis	PA	15108	Envirotech Corporation
Bordrick	Duane	9420 Telstar Avenue	El Monte	CA	91731	South Coast Air Qual. Hgmt. Distr.
Borgwardt	Robert H.	MD-65	Res. Tri. Park	NC	27111	U.S. EPA, - IERI-RTP
Borio	D. C.	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering Inc.
Borsare	David	1000 Prospect Hill	Windsor	CT	06095	Combustion Engineering
Bowen	George	85 Research Road	Hingham	MA	02043	Martek Inc.
Boyling	Chester H.	515 S. Flower St., Ste. 863	Los Angeles	CA	90071	Atlantic Richfield
Boyd	D.	P. O. Box 87	Knoxville	TN	37901	Carborundum Company
Boyd	Robert B.	6200 Oak Tree Boulevard	Independence	OH	44130	Arthur G. McKee & Company
Braden	Herbert H.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Brandt	Robert C.	P. O. Box 47837	Atlanta	GA	30362	The Cadre Corporation
Breunig	Bill	2000 Market Street	Philadelphia	PA	19103	FMC-ALKALI
Briggs	H. R. "Pat"	1 River Road	Schenectady	NY	12345	General Electric
Broadus	W. R.	P. O. Box 1603	Dalton	GA	30720	Dalton Rock Products Company
Brown	Charles S.	161 E. 42nd Street	New York	NY	10017	Koch Engineering Co., Inc.
Brown	Ralph	1129 Bellwood Avenue	Bellwood	IL	60104	Faville-LeVally Corporation
Brown	W. S.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Company
Browne	M. B.	One Oliver Plaza	Pittsburgh	PA	15222	Dravo Corporation
Broz	Larry D.	3203 Womens Club Drive, Suite 220	Raleigh	NC	27612	Acurex Corporation
Bruce	R. R.	Mississauga	Ontario	CANADA	LSK IB3	Ontario Research Foundation
Bryson	George H.	P. O. Box 908	Bloomington	IN	47401	Hoosier Energy Division
Bryson	Kevin R.	3 Executive Campus	Cherry Hill	NJ	08034	Stone & Webster Eng. Corp.
Buckingham	Paul A.	3333 Michelson Drive	Irvine	CA	92715	Fluor E & C
Budin	Michael	Brook Road	Conshohocken	PA	19428	Trane Thermal Company
Bumpass	Tom	P. O. Box 7444	Longview	TX	75602	Texas Eastman Company
Burbank	Devey A.	50 Beale Street	San Francisco	CA	94119	Bechtel National Inc.
Burford	David P.	800 Shades Creek Parkway	Birmingham	AL	35209	Southern Company Services
Burgess	Robert J.	P. O. Box 529100	Miami	FL	33169	Florida Power & Light
Burns	Eugene A.	P. O. Box 1620	La Jolla	CA	92038	Systems, Science & Software
Busko	William D.	P. O. Box 1975	Baltimore	MD	21203	Eastern Stainless Steel Company
Russell	Harvey	Bldg. 23, Rm. 352, 1 River Road	Schenectady	NY	12345	General Electric Company
Buzak	Jan	1111 Mo. 19th Street	Arlington	VA	22209	Energy and Environmental Analysis, Inc.
Byrne	John M.	200 West Monroe Street	Chicago	IL	60606	Fluor Power Services, Inc.
Cameron	Sidney M.	P. O. Box 2389	Knoxville	TN	37901	American Limestone Company

Campbell	Ivor E.	150 E. Broad Street	Columbus	OH	43215	Clyde Williams & Company
Canterbury	John A.	1800 FHC Drive, West	Itasca	IL	60143	FHC Corporation
Capp	Joseph A.	555 Madison Avenue	New York	NY	10022	Combustion Equipment Associates, Inc.
Carden	W.	420 Rouser Road	Corapolis	PA	15108	Envirotech Corporation
Carnow	Bertram W.	2121 W. Taylor	Chicago	IL	60612	University of Illinois Medical Center
Carpenter	John K.	P. O. Box 149	St. Louis	MO	63166	Union Electric Company
Casada	Donald A.	ORNL Box Y, 9104-1	Oak Ridge	TN	37830	Oak Ridge National Labs.
Cascy	John W.	P. O. Box 1380	Houston	TX	77001	Shell Dev. Company
Cavanagh	J. Gordon	P. O. Draver 5000	Lakeland	FL	38303	Davy Powergas, Inc.
Caylor	F.	434 Allgheny River Blvd.	Oakmont	PA	15139	Kaiser Engineers Power Corporation
Cesped	Ric	300 Lakeside Drive	Oakland	CA	94612	Teknekron
Chapman	Richard A.	2118 Milvia Street	Berkeley	CA	94704	Fluor E & C Inc.
Chatham	Russell L.	P. O. Box 35000	Houston	TX	77035	US EPA, IERL-RTP
Chatlyne	C. J.	MD-61	Res. Tri. Park	NC	27711	Envirotecring, Inc.
Cheng	Gregory H.	4233 N. United Parkway	Schiller Park	IL	60176	Ralph M. Parsons
Cherry	Millard W.	100 W. Walnut Street	Pasadena	CA	91006	Envirotech/Chemico
Chopra	P.	One Penn Plaza	New York	NY	10001	Teknekron, Inc.
Christman	Roger C.	4701 Sangamore Road	Washington	DC	20016	Brown & Root, Inc.
Chu	Richard R.	P. O. Box 3	Houston	TX	77001	Research-Cottrell, Inc.
Clark	Louise B.	P. O. Box 10087	Palo Alto	CA	94303	Chiyoda International Corporation
Clasen	Donald D.	1300 Park Place Building	Seattle	WA	98101	Seminole Electric Cooperative, Inc.
Claussen	Robert W.	2410 East Busch Boulevard	Tampa	FL	33612	IU Conversion Systems, Inc.
Cleveland	Lee	115 Gibraltar Road	Horsham	PA	19044	Engineering-Science
Cline	John R.	7903 Westpark Drive	McLean	VA	22102	Joy Manufacturing Company
Coe, Jr.	E. L.	P. O. Box 2744	Los Angeles	CA	90051	Burns & McDonnell Engineering Co.
Collier	Clark W.	P. O. Box 173	Kansas City	MO	64141	Tri-State G&T
Colyn	Clarence A.	12076 Grant Street	Thornton	CO	80241	Pullman Power Products
Conklin	Edwin R.	666 Fifth Avenue	New York	NY	10019	U.S. EPA-IERL
Cook	Judith E.	MD-60	Res. Tri. Park	NC	27711	Ill. State-Inst Natural Resources
Cooley	William J.	309 W. Washington Street	Chicago	IL	60657	R. W. Beck and Associates
Couppis	Evis C.	400 Prudential Plaza	Denver	CO	80265	Hartek Inc.
Coven	Robert	85 Research Road	Hingham	MA	02043	GPU Service Corporation
Crandall	W. A.	260 Cherry Hill Road	Parsippany	NJ	07054	Peabody Process Systems Inc.
Cranston	George	835 Hope Street	Stamford	CT	06909	Davy Powergas Inc.
Craton	Robert E.	6161 Savoy	Houston	TX	77036	U.S. Bureau of Mines
Crocker	Laird	St. Joe Zinc Co. Power Station	Monaca	PA	15061	U.S. EPA
Croze	Robert	26 W. St. Clair Street	Cincinnati	OH	45268	Royal Schelde
Crucq	C.	Vlissingen	NETHERLANDS			Pullman Kellogg Company
Cubisino	Art	433 Hackensack Avenue	Hackensack	NJ	07950	Teknekron
Cukor	Peter	2118 Milvia Street	Berkeley	CA	94704	Wheelabrator-Frye Inc., APCD
Culhane	F. R.	600 Grant Street	Pittsburgh	PA	15219	Linden Equipment Company
Cwik	R. A.	5618 W. Montrose Avenue	Chicago	IL	60634	Neptune AirPOL
Daggett	Harry	23480 Park Sorrento	Calabasas	CA	91304	American Air Filter Company
Dahlem	F. E.	215 Central Avenue	Louisville	KY	40277	Electric Power Research Institute
Dalton	Stuart M.	3412 Hillview Avenue	Palo Alto	CA	94303	Koppers Company, Inc.
Davis	Richard L.	2701 Koppers Building	Pittsburgh	PA	15219	Combustion Equipment Associates Inc.
Davis	Robert	555 Madison Avenue	New York	NY	10022	U.S. Steel Corporation
Davis	S. A.	600 Grant Street	Pittsburgh	PA	15230	Tucson Gas & Electric
De Hasi	Cosimo	220 W. Sixth Street	Tucson	AR	85702	Habcock & Wilcox Company
De Priest	William	20 S. Van Buren Avenue	Barborton	OH	44203	

Deacon	J.	420 Rouser Road	Corapolis	PA	15108	Envirotech Corporation
Demsey	J. Herbert	P. O. Box 12796	Res. Tri. Park	NC	27709	Acurex Corporation
Dene	Charles E.	P. O. Box 2000	Paducah	KY	42001	EPRI - Shawnee Test Facility
Devitt	Timothy W.	11499 Chester Road	Cincinnati	OH	45242	PEDCO Environmental, Inc.
Dharmarajan	M. N.	200 N. 7th Street	Lebanon	PA	17042	Buell ECD-Envirotech
Di Gioia, Jr.	A. H.	570 Beatty Road	Hontoeville	PA	15146	GAI Consultants Inc.
Dickerman	Jim	8500 Shoal Creek Blvd., Box 9948	Austin	TX	78766	Radian Corporation
Dietrich	David C.	100 N.E. Adams Street	Peoria	IL	61629	Caterpillar Tractor Company
Dixon	Gary N.	8900 DeSoto	Washington	DC	20460	U.S. EPA/Office of Solid Waste
Dollmeyer	Chester V.	14000 Georgia Avenue	Canoga Park	CA	91364	Rockwell International-Energy Systems Cp
Donahue	Kathleen E.	1200 N.W. 63rd, Box 10400	Silver Spring	MD	20910	Automation Industries, Inc.
Donoghoe	Clarence H.	P. O. Box 4005	Oklahoma City	OK	73156	Benham-Holway Power Group
Doody	Bernard A.	33 Industry Avenue	Champaign	IL	61820	U.S. Army, CERL
Doty	Calvin N.	P. O. Box 36444	Houston	TX	77036	Davy Powergas, Inc.
Doty	Robert W.	600 Grant St., 42nd Floor	Pittsburgh	PA	15219	Eckert, Seamans, Cherin & Mellott
Dowdy	William S.	1501 Alcoa Building	Pittsburgh	PA	15227	Aluminum Company of America
Downs	David A.	799 N. Main Street	Akron	OH	44310	Brad Associates
Dragos	William J.	1562 Beeson Street	Alliance	OH	44601	Babcock & Wilcox Company
Duffy	John	Speed Scientific School	Louisville	KY	40222	University of Louisville
Dunkle	Ron	Consol Plaza	Pittsburgh	PA	15241	Consolidation Coal Company
Durkee	David	835 Hope Street	Pittsburgh	PA	06909	Peabody Process Systems, Inc.
Dutton	Kenneth R.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Eaton	Roger W.	MD-13	Res. Tri. Park	NC	27711	U.S. EPA-ESED
Eaton	C. B.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch, Consulting Engineers
Ebrey	Terry J.	P. O. Box 211	Lakeland	FL	33803	Davy Powergas, Inc.
Ebzyer	John H.	601 Jefferson, 27th Floor	Lafayette	LA	66040	Kansas City Power & Light
Efimenko	Joan	1231 25th N.W., Bur. Nat'l Affairs	Houston	TX	77005	Lodge-Cottrell Operations Div. Dresser Ind
Eisenlohr, Sr.	Alex	1740 W. Adams St.-Bur. Air Quality	Washington	DC	20037	Environment Reporter
Elder	D. H.	Two Rector	Phoenix	AZ	85007	Arizona State Health Department
Elliott	Henry W.	OSWIA	New York	NY	10006	Ebasco Services Inc.
Engck	David	480 University Avenue	Muscle Shoals	AL	35660	Tennessee Valley Authority
Engdahl	T. W.	10 South Riverside Plaza	Toronto, Ontario	CANADA	MSG 1V2	Acres Davy Limited
Eriksen	Richard	505 King Avenue	Chicago	IL	60606	Arthur G. McKee & Company
Erskine	Robert L.	1717 E. Interstate Avenue	Columbus	OH	43201	Battelle
Esche	George	1820 Dole Madison Avenue	Bismarck	ND	58501	Basin Electric Power Cooperative
Fackler	Michael	Sulthachstr. 22, 6600	HcLean	VA	22102	The HITRE Corporation, Metrek Div.
Fail	Robert H.	818 Kansas Avenue	Saarbrücken	GERMANY	66612	Saarberg-Holter Umwelttechnik GmbH
Farber	Albert B.	P. O. Box 476	Topeka	KS	66612	Kansas Power and Light Company
Farrington	Paul S.	9700 South Cass Avenue	Beaver	PA	15009	M-K National Corporation
Favell	James	1500 E. Putnam Avenue	Argonne	IL	60439	Argonne National Laboratory
Fatherman	M. A.	Box 12121, 555 West Hastings Street	Old Greenwich	CN	06870	Flakt
Feuerborn	Donald	P. O. Box 1500	Vancouver	BC	V6B 4T6	British Col. Hydro & Power Authority
Fife	Dale F.	1330 Baltimore Street	Somerville	NJ	08876	Research Cottrell
Fike	C. H.	29 S. LaSalle Street	Kansas City	MO	64141	Kansas City Power & Light Company
Finer	Waldo E.	799 N. Main Street	Chicago	IL	60603	Babcock & Wilcox Company
Finkel	J.	420 Rouser Road	Akron	OH	44310	Brad Associates
Fish	Lawrence H.	3 Girard Plaza	Corapolis	PA	15108	Envirotech
	William H.	P. O. Box 499	Philadelphia	PA	19102	Suntech, Inc.
			Columbus	ME	68601	Nebraska Public Power District

Fisher	Ray W.	Energy & Min. Res. Rsch. Inst.	Ames	IA	50011	Iowa State University
Fitch	Robert E.	P. O. Box 3707, Mail Stop 9A-01	Seattle	WA	98124	Boeing Eng'g. and Construction Co.
Fletcher	J. C.	9041 Executive Park Drive	Knoxville	TN	37919	United Engineers & Constructors
Fling	Richards B.	2805 Vista Mesa Drive	Rancho PalosVerdes	CA	90274	The Aerospace Corporation
Foley	Gerry F.	185 Crossways Park Drive	Woodbury	NY	11797	Burns & Roe, Inc.
Forck	B. A.	Klinkenstr. 27-31, D4300	Essen	GERMANY		VGB Tech Ver. der Grosskraftwerks E.V.
Forrest	J. A.	283 Rt. 17 South	Paramus	NJ	07652	Burns & Roe, Inc.
Fowler	Carolyn P.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA-IERL
Fox	Harvey	Box 750	Somerville	NJ	08805	Research-Cottrell, Inc.
Fox	Landon D.	400 Commerce Avenue W10B104	Knoxville	TN	37902	Tennessee Valley Authority
Francis	Daniel V.	Frankfort Road	Monaca	PA	15061	ARCO/Polymers
Frank	Roger J.	P. O. Box 14219	Phoenix	AZ	85063	Hogan Manufacturing, Inc.
Franklin	Alan Lynn	P. O. Box 999	Richland	WA	99352	Battelle Northwest
Franklin	Robert C.	650 Smithfield	Pittsburgh	PA	15222	Dravo Lime Company
Frees	Joseph L.	393 Seventh Avenue	New York	NY	10001	Gibbs & Hill, Inc.
Freire	Donald E.	Box A	Monaca	PA	15061	St. Joe Minerals Corporation
Freshcorn	G. E.	5001 W. 80th Street	Minneapolis	MN	55437	Combustion Engineering, Inc.
Friedrichs	Michael J.	576 Standard Avenue	Richmond	CA	94802	Chevron Research Company
Fuchs	David A.	P. O. Box 12194	Res. Tri. Park	NC	27709	Research Triangle Institute
Fuller	D.	200 N. 7th Street	Lebanon	PA	17042	Envirotech/Buell
Furlong	Robert	The American Road, Room 626	Dearborn	MI	48121	Ford Motor Company
Futryk	Joseph L.	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer, Inc.
Gaines	T.	One Penn Plaza	New York	NY	10001	Chemico Air Poll. Control
Gallagher	Edward E.	P. O. Box 960	Cincinnati	OH	45201	Cincinnati Gas & Electric Company
Galloway	Marsh J.	140 Sheldon Road	Berea	OH	44017	The Ceilcote Company
Galloway	P.	One Penn Plaza	New York	NY	10001	Envirotech/Chemico
Gambarani	Alberto	74 Inverness Drive East	Englewood	CA	80112	Lear Siegler, Inc. Environ. Tech. Div.
Garcia	Jim R.	P. O. Box 2624	Birmingham	AL	35202	Southern Company Services
Garner	Bernie	433 Hackensack Avenue	Hackensack	NJ	07950	Pullman Kellogg
Garvin	W. D.	P. O. Box 30013	Raleigh	NC	27612	Martin Marietta Aggregates
Gatchouse	C.	420 Rouser Road	Coraopolis	PA	15108	Envirotech Corporation
Gaudette	Paul R.	6701 W. 64th Street	Shawnee Mission	KS	66202	Research-Cottrell
Gault	John	5916 Fannin Street	Houston	TX	77004	Bovay Engineers Inc.
Gaulter	Lee S.	P. O. Box 538	Allentown	PA	18105	Air Products & Chemicals Inc.
Gehri	Dennis C.	8900 De Soto Avenue	Canoga Park	CA	91304	Energy Systems Group-Rockwell International
Gilbert	Richard A.	128 Spring Street	Lexington	MA	02173	Kennecott Copper Corporation
Gilbert	Carl A.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Company
Gilmore	Donald B.	P. O. Box 15027	Las Vegas	NV	89114	EHSI-LV, HSA
Giovannetti	Albert	P. O. Drawer 5000	Lakeland	FL	33803	Davy Powergas Inc.
Giovanni	Dan V.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Glauser	John	6161 Sandy Drive	Houston	TX	77036	Davy Powergas
Gleason	Robert J.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Gleason	Thomas	470 Park Avenue, South	New York	NY	10016	Swemco Inc.
Glenn	Roland D.	50 East 41st Street	New York	NY	10017	Combustion Processes, Inc.
Gluck	Ted	7760 W. Devon Avenue	Chicago	IL	60631	EPCO
Goerner	Roger K.	576 Standard Avenue	Richmond	VA	94804	Chevron Research Company
Gogineni	H. R.	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering
Golden	Dean	3412 Hillview Drive	Palo Alto	CA	94303	Electric Power Research Institute
Goldschmidt	Klaus	Bismarckstr. 54	Essen	GERMANY	D-4300	STEAG Aktiengesellschaft
Goodwin	R. W.	P. O. Box 10087	Palo Alto	CA	94303	Research-Cottrell, Inc.

Gosik	Robert	Denver Federal Center	CO	Room	U.S. EPA
Gottlieb	Myron	ECT Division	DC	20545	Department of Energy
Gottschalk	Chris	P. O. Box 2900, Shawnee Test Fac.	KY	42001	Tennessee Valley Authority
Granda	M. R.	P. O. Box 3	TX	77001	Brown & Root Inc.
Grant	Richard J.	607 E. Adams Street, Rm. 721	IL	62701	Central Ill. Public Service Company
Grassel	Eugene E.	P. O. Box 1299	IN	55440	Donaldson Company, Inc.
Green	Clois L.	P. O. Box 472	TX	76567	Alcoa
Green	Robert H.	P. O. Box 880	TX	75001	Metal Components
Greene	Jack H.	HD-60	NC	27711	U.S. EPA
Grim	Carlton D.	40 East Broadway	MT	59701	Montana Power Company
Grim	Richard P.	4500 Cherry Creek Drive	CO	80217	Stearns-Roger, Inc.
Groves	Kenneth O.	2020 Bldg., Abbott Road	MI	48640	Dow Chemical Company
Grubb	Theron	525 S. Mayden Road	AZ	85282	Fabric Filters
Grumbrecht	Volker	P. O. Box 1949/1960 - 5270	W. GERMANY		L. & C. Steinmuller
Guay	Bill	44 Briar Ridgeroad	CT	06810	Newmont Exploration
Gude	Klaus	9165 Rumsey Road	MD	21045	Niro Atomizer Inc.
Guernsey	Edwin O.		NJ	08066	Mobil Research & Development Corp.
Gullett	David E.	100 Summer Street	MA	02110	United Engineers and Constructors
Gupta	Pat	2 Country View Road	PA	19355	Ecolaire Systems, Inc.
Gylfe	J. Donald	890 DeSoto Avenue	CA	91304	Rockwell International
Hall	John	175 E. 5th Street	IN	55101	Burlington Northern Inc.
Hall	William G.	Walden Avenue	NY		ANDCO
Halpern	Mark	1900 Pennsylvania Avenue	DC	20068	Potomac Electric Power Company
Hammersan	J. Warren	1 Space Park	CA	90278	TRW Systems and Energy
Hansen	Jim	1000 S. Fremont Avenue	CA	91802	CF Braun
Hansen	Svend Kris	9165 Rumsey Road	CA	21045	Niro Atomizer
Hargrove	Buddy	8500 Shoal Creek Blvd., Box 9948	TX	78766	Radian Corporation
Harris, Jr.	William B.	Bldg. 2, Rm. 132B1, River Road	NY	12345	General Electric Company
Harrison	Doug	800 Kipling Avenue	CANADA	H82 5S4	Ontario Hydro
Hartman	John R.	640 S. Main Street	CA	91790	Wallace & Tiernan
Hartman	Scott	11499 Chester Road	OH	45246	PEDCO Environmental, Inc.
Havlik	Frank A.	P. O. Box K-7	VA	23288	Infilco Degremont, Inc.
Hayashida	Paul K.	P. O. Box 400	CO	80201	Mitsubishi International Corporation
Hayes, Jr.	Richard D.	1501 Alcoa Building	PA	15219	Alcoa
Heacock, Jr.	Frank A.	P. O. Box 2166	AZ	85036	Arizona Public Service Company
Head	Harlan M.	50 Beale Street	CA	94119	Bechtel National Inc.
Headley	Larry C.	P. O. Box 880	WV	26505	U.S. Department of Energy
Healey	Joseph R.	835 Hope Street	CT	06909	Peabody Process Systems, Inc.
Hein	Glen	300 W. Washington	IL	60606	Marblehead Lime Company
Heinz	Elwood	433 Hackensack Avenue	NJ	07645	Pullman Kellogg
Heinze	Kay	420 Rouser Road	PA	15108	Envirotech
Henderson	W. B.	200 N. 7th Street	PA	17042	Envirotech/Buell
Hendron	David H.	55 E. Monroe Street	IL	60044	Woodward-Clyde Consultants
Hennico	A.	Boite Postale 311	FRANCE	92506	Institut Francais du Petrole
Hentges	Robert A.	6105 Center Hill Road	OH	45224	Procter & Gamble Company
Herlihy	Jim	401 M Street, S.W.	DC	20460	U.S. EPA
Hess	H. F.	1800 FHC Drive, West	IL	60143	FMC Corporation
Hettler	Robert F.	2800 Mitchell Drive	CA	94598	Dow Chemical Company
Hickok	Wayne	3316 W. 66th Street	IN	55435	Cooperative Power Association
Hill	David Lee	44 Briar Ridge Road	CT	06810	Newmont Exploration Limited

Hill	Russell Ward	P. O. Box 908	Bloomington	IN	47401	Hoosier Energy Division
Hilton	Robert	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Hintz	Charles F.	P. O. Box 2825	Allentown	PA		Mosser/Ecolaire
Hoffman	Frank W.	P. O. Box 269	Springfield	MA	01101	Calfran Industries
Hojnicki	Tom	9700 Argonne National Labs.	Argonne	IL	60439	Argonne National Labs.
Hollett, Jr.	Grant T.	78015 York Road	Bensenville	IL	60106	Flick-Reedy Corporation
Hollinden	G. A.	470 CUBB	Chattanooga	TN	37401	Tennessee Valley Authority
Hong	Sun-nan	P. O. Box 538	Allentown	PA	18105	Air Products & Chemicals, Inc.
Hooker	Neal	4282	N. Canton	OH	44720	Babcock & Wilcox
Horwand	Svend	Rumsey Road	Columbia	MD	21045	Niro Atomizer
Horn	Nelson E.	2100 Clearwater Drive	Oak Brook	IL	60540	Combustion Engineering Inc.
Horn	Richard J.	Two Countryview Road	Malvern	PA	19355	Ecolaire Systems Inc.
Horowitz	A. L.	6105 Center Hill Road	Cincinnati	OH	45224	Procter & Gamble
Howitz	H.	One Penn Plaza	New York	NY	10001	Envirotech/Chemico
Rosemann	Robert P.	77 Beale Street	San Francisco	CA	94106	Pacific Gas & Electric Co.
Howell	R. D.	100 W. Walnut Street	Pasadena	CA	91124	Ralph M. Parsons Company
Hoynes	T.	420 Rouser Road	Coraopolis	PA	15108	Envirotech Corporation
Hsieh	C. K.	P. O. Box 800	Rosemead	CA	91770	Allegheny Ludlum Steel Corporation
Hunt	Joseph H.	Research Center	Brackenridge	PA	15014	Allied Chemical Corporation
Hunter	W. D.	P. O. Box 1139R	Morristown	NJ	07960	Babcock & Wilcox Company
Hurst	T. B.	20 S. Van Buren Avenue	Barberton	OH	44203	Chemico Air Pollution Control Corp.
Hurwitz	Howard	One Penn Plaza	New York	NY	10001	Andco Power Industry Products, Inc.
Hushen	Thomas H.	1880 Republic Avenue	San Leandro	CA	94577	Betz Labs. Inc.
Hutta	Paul J.	Somerton Road	Trevose	PA	19047	Komline-Sanderson
Iannelli	Eric A.	100 Holland Avenue	Pcapack	NJ	07977	Gilbert Associates, Inc.
Ingram	T. J.	P. O. Box 1498	Reading	PA	19603	Stearns Roger
Ireland	Paul	P. O. Box 5888	Denver	CO	80217	Chemico Air Poll. Control
Ivask	S.	One Penn Plaza	New York	NY	10001	Exxon Research & Engineering Co.
Jahnig	Charles E.	P. O. Box 8	Linden	NJ	07036	Proto-Power Management Corporation
Jalbert	Paul E.	591 Poquonock Road	Groton	CT	06340	Basin Electric Power Cooperative
Janssen	Kent E.	1717 E. Interstate Avenue	Bismarck	ND	58501	Martin Marietta Corp.
Jashnani	I.	1450 S. Rolling Road	Baltimore	MD	21227	Foster Wheeler Energy Corporation
Jensen	Alan	105 W. Adams	Chicago	IL	60603	Bechtel Power Corporation
Jensen	Robert H.	50 Beale Street	San Francisco	CA	94105	Pullman Power Products
Jewell	Robert G.	1700 West Loop South, Suite 1145	Houston	TX	77027	Burns & McDonnell
Johannes	Richard D.	P. O. Box 173	Kansas City	MO	64141	Peabody Process Systems, Inc.
Johnson	Carlton	835 Hope Street	Stamford	CT	06909	C. F. Braun & Company
Johnson	Howard	361 East Broad Street	Columbus	OH	43212	Exxon Research & Engineering
Johnson	Robert A.	1000 Fremont Avenue	Alhambra	CA	91802	Gifford-Hill & Company, Inc.
Johnson	Russell R.	P. O. Box 101	Florham Park	NJ	07932	SRI International
Jones	J. R.	P. O. Box 47127	Dallas	TX	75247	U.S. EPA-IERL
Jones	Jerry L.	333 Ravenswood Avenue	Menlo Park	CA	94025	Hoosier Energy Division
Jones	Julian W.	MD-61	Res. Tri. Park	NC	27711	Air Products & Chemicals Inc.
Jones	Marion	P. O. Box 908	Bloomington	IN	47401	Research-Cottrell, Inc.
Jones	Robert D.	P. O. Box 538	Allentown	PA	18105	Envirotech
Judersleben	Peter	P. O. Box 1600	Somerville	NJ	08876	New York Power Pool
Jurgensen	H. D.	400 Rouser Road	Coraopolis	PA	15108	Chiyoda International Corporation
Juris	Kenneth	3890 Carman Road	Schenectady	NY	12303	Texas Air Control
Kamcoka	Y.	1300 Park Place Building	Seattle	WA	98101	
Kamrath	Jim	1300 Larkspur	Austin	TX	78758	

Kang	Cecilia C.	450 East Edsall Boulevard	Palisades Park	NJ	07650	Catalysis Research Corporation
Kaplan	Marilyn	2970 Maria Avenue	Northbrook	IL	60062	McIlvaine Company
Kaplan	Norman	MD-61	Res. Tri. Park	NC	27711	U.S. EPA, IERL-RTP
Kasik	Steven M.	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer Inc.
Kaye	Lawrence A.	2000 Second Avenue	Detroit	MI	48226	Detroit Edison Company
Keen	Ron	835 Hope Street	Stamford	CT	06909	Peabody Process Systems Inc.
Kendle	R. T.	P. O. Box 140232	Charlotte	NC	28224	Catalytic, Inc.
Kent	James R.	1800 FHC Drive, West	Itasca	IL	60143	FHC Corporation
Kesler	Raymond	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell
Kesler	Rick	P. O. Box 16067	Denver	CO	80216	Hine and Smelter
Kesstner	Mark	35 South Jefferson Road	Whippany	NJ	07981	Apollo Chemical Corp.
Kettner	Don	Power Production	Elk River	MN	5330	United Power Association
Kiff	John	4565 Coco Boulevard	Los Angeles	CA	90039	Joy Manufacturing (WP Div.)
Kilgroe	James D.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA-IERL
Kincaid	Jim	6100 Center Hill Road	Cincinnati	OH	45231	Procter & Gamble
Kindig	James K.	4601 Indiana Street	Golden	CO	80401	Hazen Research, Inc.
Klemovich	Ronald H.	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell
Knefelkamp	Robert A.	3800 Race Street	Denver	CO	80216	Mine & Smelter Corporation
Knight	R. Gordon	4301 Dutch Ridge Road	Beaver	PA	15009	Michael Baker, Jr., Inc.
Knowles	W. D.	P. O. Box 3105	Houston	TX	77001	Shell Oil Company
Koba	S.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Kodras	Frank D.	1500 Market Street	Philadelphia	PA	19102	Catalytic, Inc.
Kolbe	John L.	1500 S. Rolling Road	Baltimore	MD	21227	Martin Marietta Labs.
Kondo	Ken	925 S. Niagara Street	Denver	CO	80224	Mitsubishi International Corporation
Konig	Robert J.	P. O. Box 300	Tulsa	OK	74102	Cities Service Company
Korman	Samuel	632 W. 125 Street	New York	NY	10027	Columbia University
Koshkin	H.	One Penn Plaza	New York	NY	10001	Chemico Air Poll. Control
Kovach	John J.	Collins Ferry Road	Morgantown	WV	16505	Department of Energy
Koval	T.	P. O. Box 87	Knoxville	TN	37901	Carborundum Company
Krause	Robert E.	955 Mcarns Road	Warminster	PA	18974	Pennwalt Corp.-Sharples-Stokes Div.
Krekels	Y. T.	57 Engelen Kampstraat	Sittard 6131 J.E.	NETHERLANDS		NEOM B.V.
Kreuziger	Wolfgang	Stauffenbergstrasse 26,	Berlin W 30	W. GERMANY		Bewag-Berliner Kraft-und Licht AG
Kronenberger	Robert W.	20 S. Van Buren	Barberton	OH	44203	Babcock & Wilcox Company
Krumme	J. Lee	2555 Cumberland Parkway, N.W.	Atlanta	GA	30339	Vinings Chemical Company
Kuo	Wen L.	12400 E. Imperial Highway	Norwalk	CA	90650	Bechtel
Kuzela	Ed	24 Perimeter Center East	Atlanta	GA	30348	UOP Air Correction Division
Lagaras	John S.	P. O. Box 23210	Oakland	CA	94623	Kaiser Engineers, Inc.
Lamb	Jack F.	P. O. Box 189	Addison	TX	76001	Forney Engineering Company
Lambert	Steve L.	920 S. W. 6th Avenue	Portland	OR	97204	Pacific Power & Light Company
Langeland	Wes	85 Research Road	Hingham	MA	02043	Martek Inc.
Lanois	G. D.	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Laseke	Bernie A.	11499 Chester Road, Chester Tower	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Lasky, Jr.	Bernard A.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Laslo	Dennis	P. O. Box 1107	Darien	CT	06840	Air Correction Div., UOP
Laughlin	James H.	P. O. Box 12796	Res. Tri. Park	NC	27709	Acurex Corporation
Lawson	Clifton	P. O. Box 9538	Madison	WI	53579	Warzyn Engineering, Inc.
Lawson	P.	700 University Avenue EC M5G1X6	Toronto	CANADA		Ontario Hydro
Layman	George O.	P. O. Box 1151	Pensacola	FL	32520	Gulf Power Company
LeBoeuf	Gene	3913 Algoma Road	Green Bay	WI	54301	Feeco International Inc.
Lee	George C.Y.	50 Beale Street	San Francisco	CA	94119	Bechtel National Inc.

Lee	John R.	P. O. Box 1010	Tuscaloosa	AL	35401	B. F. Goodrich Engineered Systems Co.
Leffmann	Warren W.	6120 Westover Drive	Oakland	CA	94611	Bechtel Power Corporation
Legatski	L. Karl	FHC Drive, West	Itasca	IL	60143	FHC Corporation
Legier	William B.	Prudential Center	Boston	MA	02199	Charles T. Main, Inc.
Lehtola	Philip R.	29801 Euclid Avenue	Wickliffe	OH	44121	Bailey Controls Company
Leitman	J. D.	P. O. Box 87	Knoxville	TN	37901	Carborundum Company
Leivo	Charles	P. O. Box 19566	Irvine	CA	92713	Dresser Industries
Levis	James J.	2200 Churchill Street	Springfield	IL	62706	Illinois Environ. Prot. Agency
Levy, Jr.	Newton	2 Executive Plaza	Hunt Valley	MD	21030	Martin Marietta Chemical
Lewis	C. J.	P. O. Box 15453	Lakewood	CO	80215	National Lime Association
Lingois	William A.	Stanley Building	Muscatine	IA	52761	Stanley Consultants
Lillestolen	Tom	1500 E. Putnam	Old Greenwich	CN	06870	Flakt Inc.
Lingle	Jim	231 W. Michigan Street	Milwaukee	WI	53201	Wisconsin Electric Power Company
Link	F. William	591 Marquette Mall	Michigan City	IN	46360	Northern Indiana Public Service Co.
Lisewski	Alfred A.	Two Country View Road	Malvern	PA	19355	Ecolaire Systems Inc.
Lisk	Ian	1301 South Grove	Barrington	IL	60025	Technical Publishing Company
Lloyd	Charles G.	1415 Rollins Road	Burlingame	CA	94010	Sharples-Stokes Div., Pennwalt Corp.
Long	H. E.	600 Grant Street	Pittsburgh	PA	15219	Rockwell International/Joint Venture
Longfellow	R. L.	257 Geneva Drive	Aliquippa	PA	15001	Dravo Lime Company
Loquercio	Peter	309 West Washington	Chicago	IL	60606	Ill. Institute Natural Resources
Lorring	Rick	51 Corporate Woods	Overland Park	KS	66210	Olin Water Services
Lowell	Philip S.	4107 Medical Parkway, #214	Austin	TX	78756	P. S. Lowell & Co., Inc.
Luce	Ken	4565 Colorado Blvd.	Los Angeles	CA	91006	Joy Manufacturing Company
Lucy	Thomas H.	P. O. Box 1500	Somerville	MA	08876	Research-Cottrell
Lundy	Terry	P. O. Box 230	Las Vegas	NV	89151	Nevada Power Company
Lunt	Richard R.	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Lutz	Stephen J.	201 N. Roxboro	Durham	NC	27701	TRW Environmental Engineering
Macaskill	D.	55 Avenida De Orinda	Orinda	CA	94563	Macaskill Associates
MacDonald	Bruce	85 Research Road	Hingham	MA	02043	Martek Inc.
MacRae	Terry	Two Country View Road	Malvern	PA	19355	Ecolaire Systems, Inc.
Maddalone	Ray	One Space Park	Redondo Beach	CA	90278	TRW Inc.
Madenburg	Richard S.	P. O. Box 7808	Boise	ID	83729	Morrison-Knudsen Company, Inc.
Madonia	John J.	55 E. Monroe	Chicago	IL	60603	Sargent & Lundy
Magtoto	Artemio R.	P. O. Box 1139R	Morristown	NJ	07960	Allied Chemical
Hajdeski	H. M.	20 So. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Company
Hakar	John E.	1800 FHC Drive, West	Itasca	IL	60143	FHC Corporation
Maley	L. B.	4550 W. 109th Street	Overland Park	KS	66210	Babcock & Wilcox
Malik	Kal	1000 Prospect Hill	Windsor	CT	06095	Combustion Engineering
Mann	Earl L.	5265 Hohman Avenue	Hammond	IN	46325	Northern Indiana Public Service Co.
Marby	R. F.	10 South Riverside Plaza	Chicago	IL	60606	Arthur G. McKee & Company
Marder	Sidney	2058 Huntleigh Road	Springfield	IL	62704	Harder and Associates
Mardirossian	Aris	P. O. Box 5691	Derwood	MD	20855	Indianapolis Power & Light Company
Martin	C. K.	P. O. Box 1595B	Indianapolis	IN	46206	Burns & McDonnell
Martin	James R.	1000 Prospect Hill Road	Kansas City	MO	04141	Combustion Engineering, Inc.
Mason	Louis	145 Cedar Lane	Windsor	CT	06095	Neptune Airpol Inc.
Mason	Thomas O.	General Motors Tech. Ctr.	Englewood	NJ	07631	General Motors Corporation
Hatoj	H. J.	P. O. Box 217	Warren	MI	48090	Kaiser Steel Corporation
Maxwell	Michael A.	HD-61	Fontana	CA	92335	U.S. EPA, IERI-RTP
Mayfield	John R.	2005 Walden Avenue	Res. Tri. Park	NC	27711	Abdco Power Industry Pro.
			Buffalo	NY	14240	

McCarthy	Carol L.	8500 Shoal Creek Blvd.	Austin	TX	78766	Radian Corporation
McGormick	C. J.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Company
McCurdy	Wayne	200 Massachusetts Avenue, N.W.	Washington	DC	20545	Department of Energy
McDowell III	Robert V.	800 King Street	Washington	DE	19899	Delmarva Power & Light Company
McFarlane	Lloyd R.	12631E Imperial Highway, Suite 107B	Wilmington	CA	90670	Ecolaire Inc.
McGlamery	Gerald G.	OSWA	Muscle Shoals	AL	35660	Tennessee Valley Authority
McIlvaine	Robert	2970 Maria	Northbrook	IL	60062	McIlvaine Company
Hedrows	Michael L.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch
Mehta	Dhirendra C.	555 Madison Avenue	New York	NY	10022	Combustion Equipment Associates, Inc.
Meinig	R. E.	555 Madison Avenue	New York	NY	10022	Combustion Equipment Associates, Inc.
Merdes	John	P. O. Box 10087	Palo Alto	CA	94303	Research-Cottrell, Inc.
Merlet	R.	P. O. Box 87	Knoxville	TN	37901	Carborundum Company
Herrill	Heinz	Gervinusstrasse 17/19 6000	Frankfurt/Main	GERMANY		Lurgi Umwelt und Chemotechnik GmbH
Messing	Richard S.	485 Clyde Avenue	Houtain View	CA	94042	Acurex Corporation
Meyers	Aubrey F.	1271 Avenue of the Americas	New York	NY	10020	Empire State Electric Energy Res. Corp.
Meyler	James E.	2000 Second Avenue	Detroit	MI	48226	The Detroit Edison Company
Michels	J. A.	4565 Colorado Blvd.	Los Angeles	CA	90039	Joy Manufacturing Company
Michener	Harold T.	1-New York Plaza	New York	NY	10004	International Nickel Co., Inc.
Miller	A. W.	Columbia Road	Morristown	NJ	07960	Allied Chemical Corporation
Miller	Bruce A.	1 Broadway	Cambridge	MA	02142	Badger America Inc.
Miller	Dick	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Miller	E. Mack	380 Civic Drive	Pleasant Hill	CA	94523	Industrial Clean Air
Miller	Michael W.	P. O. Box 24407	Ft. Lauderdale	FL	33307	Parkson Corporation
Minnella	Thomas J.	2964 LRJ Freeway	Dallas	TX	75234	IUP/Air Correction Division
Mirchandani	T. H.	300 Lakeside Drive	Oakland	CA	94623	Kaiser Engineers Power Corporation
Mobley	J. David	HD-61	Res. Tri. Pk.	NC	27711	U.S. EPA, IERL-RTP
Mohn	Nancy C.	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering Corp.
Moody	Harvey C.	111 Windsor Drive	Port Edwards	WI	54469	Nekoosa Papers Inc.
Morasky	Ron	3412 Hillview Avenue	Oak Brook	IL	60067	Sharples Centrifuges
Morelli	Thomas H.	105 South Meridian	Palo Alto	CA	94303	Electric Power Research Institute
Moser	Mark H.	P. O. Box 3822	Indianapolis	IN	46225	Amac Coal Company
Mullen	Robert	115 Gibraltar Road	San Francisco	CA	94119	Brown & Root, Inc.
Murad	Hugh	555 Madison Avenue	Horsham	PA	19044	IU Conversion Systems, Inc.
Muren	Fred Y.	1211 W. 22nd Street	New York	NY	10022	Combustion Equipment Associates, Inc.
Murphy	E. J.	Bldg. 2, Rm. 710, 1 River Road	Oak Brook	IL	60521	Research-Cottrell, Inc.
Murray	Kenneth R.	P. O. Box 1440	Schenectady	NY	12345	General Electric Company
Musgrove	Daniel N.	520 South Post Oak	Erie	PA	16533	Hammermill Paper Company
Mutsakis	John	161 East 42nd Street	Houston	TX	77027	Bechtel Power Corporation
Naeve	Michael	P. O. Box 1700	New York	NY	10017	Koch Engineering Company
Nakabayashi	Steve	Thermal Power Department	Houston	TX	77001	Houston Lighting & Power Company
Namann	Yasuyuki	300 W. Washington Street	Tokyo	JAPAN		Electric Power Development Co., Ltd.
Nelson	C. E.	607 East Adams Street	Chicago	IL	60606	Harbhead Lime Company
Ness	T. R.	P. O. Box 8213, University Station	Springfield	IL	62701	Central Ill. Public Service Company
Newby	Harvey M.	1310 Beulah Road	Grand Forks	ND	58202	Department of Energy
Newham	Richard A.	835 Hope Street	Pittsburgh	PA	15235	Westinghouse Electric Corporation
Newman	Thomas	Two Country View Road	Stanford	CT	06907	Peabody Process Systems, Inc.
Nguyen	Carl L.	P. O. Box 3	Halverne	PA	19355	Ecolaire Systems Inc.
Nicholas	Thuyet Duc	Trans. Can. Highway	Houston	TX	77001	Brown & Root, Inc.
	Xuan T.	1550 Northwest Highway	Senneville, Quebec	CANADA	HQX 3L7	Domet Inc. Research Centre
	George W.		Park Ridge	IL	60068	Dames & Moore

Nissen	William I.	1600 E. First Street	Salt Lake City	UT	84112	U.S. Bureau of Mines
Nixon	David C.	30 W. Superior Street	Duluth	MN	55802	Minnesota Power & Light Company
Nixon	W. Robert	77 Beale Street	San Francisco	CA	94106	Pacific Gas & Electric Co.
Modiff	Marvin J.	325 W. Adams Street	Springfield	IL	62706	Ill. Institute of Natural Resources
Morton	Richard D.	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell
Novack	Robert	85 Research Road	Hingham	MA	02043	Hartek Inc.
Novak	R. J.	5509 Tarrywood Court	Raleigh	NC	27609	Energy Resources
O'Brien	Joel C.	415 Power Building	Chattanooga	TN	37401	Tennessee Valley Authority
O'Brien, Jr.	William E.	OSMA	Muscle Shoals	AL	35660	Tennessee Valley Authority
Oestermeyer	A. W.	P. O. Box 1500 S	Somerville	NJ	08876	Research-Cottrell
Oguchi	Kenneth W.	4809 Todd Avenue	E. Chicago	IN	46312	Graver Energy Systems, Inc.
Okazawa	Tomoyoshi	700 S. Flower Street	Los Angeles	CA	90017	Sumitomo Metal America Inc.
Oldenkamp	K.	1 Houston Center Ste. #1810	Houston	TX	77002	Japan Trade Center
O'Leary	Richard D.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Oliver	Norman F.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Olsson	Earl D.	333 Ravenswood Avenue	Menlo Park	CA	94025	SRI International
Osohundo	Lars	1500 E. Putnam Avenue	Greenwich	CT	06830	Flakt Inc.
Ongemach	George A.	1800 FHC Drive, West	Itasca	IL	60143	FMC Corporation
Onnen	Ken	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Orcu	James H.	6702 Hollow Tree Road	Louisville	KY	40228	American Air Filter Company
Osborne	Sidney R.	700 N. Fairfax Street, Suite 304	Alexandria	VA	22314	Industrial Gas Cleaning Institute
Ostroff	Michael C.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA, IERL-RTP
Owen, Jr.	Norman	835 Hope Street	Stamford	CT	06907	Peabody Process Systems Inc.
Ozol	Hamilton S.	2600 Blair Stone Road	Tallahassee	FL	32301	Florida Dept. of Env. Regulation
Parikh	Michael A.	1450 S. Rolling Road	Baltimore	MD	21227	Martin Marietta Laboratories
Parikh	K. N.	P. O. Box 1500	Somerville	NJ	08876	Research Cottrell, Inc.
Parish	Lax	2845 Clearview Place	Atlanta	GA	33092	The CADRE Corporation
Parke	Helen R.	4614 Ramsey	Austin	TX	78750	University of Texas
Parker	John	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Parmley	C. E.	420 Rouser Road	Corapolis	PA	15108	Envirotech Corporation
Parsons	Randy D.	4614 Ramsey	Austin	TX	78756	University of Texas
Patkar	Steve	12076 Grant Street	Thornton	CO	80241	Tri-State G&T
Patten	Avi	P. O. Box 2825	Bethlehem	PA	18001	Mosser/Ecolaire
Patton	Lloyd J.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Pendergraft	Thomas W.	125 Baker Street	Costa Mesa	CA	92626	M. C. Patten & Co., Inc.
Pepper	Richard W.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Peters	Lynn K.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA-IERL
Peters	Wesley W.	P. O. Box 111	Los Angeles	CA	90051	Los Angeles Dept. of Water and Power
Peters	H. J.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye Inc., APCD
Petrie	Warren D.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA, IERL-RTP
Petrou	Jim	P. O. Box 227	Waterflow	NH	87421	Public Service Company of New Mexico
Petti	Gus	P. O. Box 880	Addison	TX	75001	Metal Components
Phelan	V.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye Inc., APCD
Phillips	John H.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye
Piasecki	James B.	P. O. Box 2180, Dresser Tower	Houston	TX	77001	The Carter Oil Company
Pickering	E. J.	General Motors Tech. Ctr.	Warren	MI	48090	General Motors Corporation
Pierce	H. C.	110 S. Orange Avenue	Livingston	NJ	07039	Foster Wheeler
Pirron	Gary G.	CNO 27	Trenton	NJ	08625	NJ Depart. of Environ. Protection
Pitcher	L. J.	10 South Riverside Plaza	Chicago	IL	60606	Arthur G. McKee & Company
	Norman D.	2 North 9th Street	Allentown	PA	18101	Pennsylvania Power & Light

Pless	Don	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Plumley	A. L.	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering, Inc.
Plyler	Everett L.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA, IERL-RTP
Plys	A. G.	400 East Sibley Blvd.	Harvey	IL	60426	ARCO Petroleum Products Co.
Ponder	Thomas	11499 Chester Road	Cincinnati	OH	45246	PEDCO Environmental, Inc.
Ponder	Wade H.	MD-61	Res. Tri. Park	NC	27711	U.S. EPA, IERL-RTP
Pope	Kenneth S.	520 S. Post Oak	Houston	TX	77001	Bechtel Power Corporation
Potterton	S. T.	P. O. Box 2423	N. Canton	OH	44720	Babcock & Wilcox Company
Powell	James H.	Stanley Building	Muscatine	IA	52761	Stanley Consultants
Preston	George T.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Princiotta	Frank T.	401 M Street, S.W.	Washington	DC	20460	U.S. EPA-OEMI
Prodesky	E. J.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Company
Provol	Steve J.	P. O. Box 1380	Houston	TX	77063	Shell Development Company
Pullen	T. K.	110 Sutter Street	San Francisco	CA	94104	Ecolaire, Inc.
Pullman	D.	200 N. 7th Street	Lebanon	PA	17042	Envirotech/Buell
Pursell	L. A.	101 S. Wacker Drive	Chicago	IL	60606	United States Gypsum Company
Quackenbush	V. C.	Centre Square West-1500 Market Street	Philadelphia	PA	19102	Catalytic, Inc.
Rabb	Dave	P. O. Box 2900, Shawnee Steam Plant	Paducah	KY	42001	Bechtel National Inc.
Raben	Irwin A.	44 Montgomery St., Suite 4220	San Francisco	CA	94104	Combustion Equipment Associates, Inc.
Ramirez	Agustin A.	1800 FMC Drive, West	Itasca	IL	60143	FMC Corporation
Rao	Richard	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell
Rautzen	Robert	P. O. Box 1123	Dayton	OH	45401	Chemineer Inc.
Ray	William G.	235 East 42nd Street	New York	NY	10017	Pfizer Inc.
Reichard	Herman	1 Moritine Plaza	San Francisco	CA	94111	Combustion Engineering, Inc.
Reid	John C.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Reilly	John B.	555 Madison Avenue	New York	NY	10022	Combustion Equipment
Reinauer	Thomas V.	10 Chatham Road	Summit	NJ	07901	Mikropul Corporation
Reisinger	A. A.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye Inc., APCD
Remillieux	Jean	19 Avenue Dubonnet	Courbevoie 92400	FRANCE		Air Industrie
Renberg	W.	420 Rouser Road	Coraopolis	PA	15108	Envirotech Corp.
Renko	Ronald	P. O. Box 372	Wellsville	NY	14895	C-E Air Preheater Co.
Retz	John A.	P. O. Box 1975	Baltimore	MD	21203	Eastern Stainless Steel Company
Reynolds	Karen Anthony	2600 Blair Stone Road	Tallahassee	FL	32301	Florida Dept. of Env. Regulation
Rhodes	Robin B.	Andover Road	Wellsville	NY	14895	CE Air Preheater
Rhudy	Richard	3412 Hillview	Palo Alto	CA	94303	Electric Power Research Institute
Ricci	Hugh	201 S. Fall Street, Capitol Complex	Carson City	NV	89710	Nevada Div. of Environmental Protection
Richardson	Phillip M.	1007 Market Street	Wilmington	DE	19898	E.I. du Pont de Nemours & Co., Inc.
Richman	Mark	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell Inc.
Richmond	Philip F.	P. O. Box 127	Center	ND	58530	Minnkota Peer Coop
Rieland	William G.	1800 Washington Road	Pittsburgh	PA	15241	Consolidation Coal Company
Riggs	Keith A.	12301 Kurland	Houston	TX	77034	Houston Lighting & Power
Ring	Leon E.	400 Commerce Avenue	Knoxville	TN	37902	Tennessee Valley Authority
Ritchie	Charles I.	2475 E. 22nd Street, Suite 510	Cleveland	OH	44115	Chemico Air Pollution Control Co.
Robbins	Stephen H.	P. O. Box 3	Houston	TX	77001	Brown & Root, Inc.
Rochelle	Gary	Dept. of Chemical Engineering	Austin	TX	78712	University of Texas
Roe	S. F.	P. O. Box 6428	Ft. Myers	FL	33904	The Hunters Corporation
Rogers	Kenneth J.	555 Madison Avenue	New York	NY	10022	Combustion Equipment
Rogers	Wyatt	7100 Broadway #3D	Denver	CO	80221	York Research
Rohlik	Ron	P. O. Box 197X, Rt. 1	Bakersfield	CA	93308	Getty Oil Company
Rollins	K. B.	P. O. Box 47320	Dallas	TX	75247	Celanese Chemical Co., Inc.

Rosenberg	Harvey S.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Division
Ross	Dennis R.	76 S. Main Street	Akron	OH	44308	Ohio Edison Company
Ross	Donald W.	P. O. Drawer 5000	Lakeland	FL	33803	Davy Powergas, Inc.
Ross	R. W.	P. O. Box 1958	Huntington	WV	25720	Huntington Alloys, Inc.
Rossoff	Jerry	2350 E. El Segundo Blvd.	El Segundo	CA	90245	Aerospace Corporation
Rubin	Edward S.	Schenley Park	Pittsburgh	PA	15213	Carnegie-Mellon University
Ruggiano	Lou	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Rukovena	Frank	P. O. Box 350	Akron	OH	44309	Norton Company/Stowe
Sainz	Darwin E.	201 S. Broadway	Orcutt	CA	93454	Union Oil Company of California
Saleem	Abdus	One Penn Plaza	New York	NY	10001	Chemico Air Pollution Control Corp.
Sannes	Carl	414 Niccollet Mall	Minneapolis	MN	55401	Arthur D. Little, Inc.
Santhanam	C. J.	20 Acorn Park	Cambridge	MA	02140	TRW Incorporated
Santy	Myrrl	One Space Park	Redondo Beach	CA	90278	Versar, Inc.
Sargent	Donald H.	6621 Electronic Drive	Springfield	VA	22151	Ecolaire Systems, Inc.
Schaffer	John H.	2 Country View Road	Halverne	PA	19355	Fluor Engineers and Constructors, Inc.
Schendel	Ronald L.	3333 Michelson Drive	Irvine	CA	92730	Carborundum Company
Scher	J.	P. O. Box 87	Knoxville	TN	37901	Enviroengineering, Inc.
Schorsch	Henry	4233 N. United Parkway	Schiller Park	IL	60276	Wheelabrator-Frye Inc., APCD
Schreyer	M. P.	14920 S. Main Street	Gardena	CA	90248	Salt River Project
Schroeder	Rick	P. O. Box 1980	Phoenix	AZ	85001	Koch Engineering Company
Schwartz	Richard	161 E. 42nd Street	New York	NY	10017	Carborundum Company
Scott	W.	P. O. Box 87	Knoxville	TN	37901	Pullman Kellogg
Scotti	Louis	433 Hackensack Avenue	Hackensack	NJ	07061	Texasgulf Inc.
Scroggins	James Edwin	P. O. Box 100	Granger	WY	82934	IU Conversion Systems, Inc.
Seabrook, Jr.	B. Lawrence	115 Gibraltar Road	Horsham	PA	19044	Lower Colorado River Authority
Scafe	William C.	Box 220	Austin	TX	78767	Selle Alloys & Equipment Co.
Selle	J. B.	1629 Bonnie Brae	Houston	TX	77006	Pfizer Inc.
Semrau	Konrad	333 Ravenswood Avenue	Henlo Park	NY	10017	Nevada Div. Environmental Protection
Senatore	Peter J.	235 East 42nd Street	New York	NY	89710	Argonne National Laboratory
Servick	Dick	Capitol Complex	Carson City	NV	60515	Wheelabrator-Frye Inc. APCD
Shaffer	Vaclav J.	2700 S. Cass Avenue	Argonne	IL	15219	Electric Power Research Institute
Shah	K. O.	600 Grant Street	Pittsburgh	PA	94303	Fischer E. Porter Company
Shah	N. D.	3412 Hillview Avenue	Palo Alto	CA	18969	Canadian Industries Limited
Shanks	Alfred T.	581 Creamery Road	Telford	PA	J3G1T9	U.S. EPA, IERL-RTP
Sharp	John A.	Chem. Res. Lab., McMasterville	Quebec	CANADA	27711	Rechtel National Inc.
Sharpe	Patricia	MD-60	Res. Tri. Park	NC	94119	Mitsubishi Heavy Industries, Ltd.
Sherwin	R. H.	50 Beale Street	San Francisco	CA	74601	Continental Oil Company
Shimizu	Taku	MCEC 118 Tomihisa-Cho	Ichigaya, Shinjuku-Ku	TKYO	85021	Arizona Public Service Company
Shimoda	Elwyn	P. O. Box 1267	Ponca City	OK	35660	SAS Corporation
Simko	Alexander P.	10025 North 21st Avenue	Phoenix	AZ	12054	NY State Energy R&D Authority
Slack	A. V.	Route 1, Box 69	Sheffield	AL	77084	Pullman Kellogg
Slaughter	Dale H.	Empire State Plaza	Albany	NY	06340	Proto-Power Management Corporation
Sliger	A. Glenn	R&D Center, Park 10	Groton	CT	07601	U.S. EPA, Region 7
Slingsby	Donald K.	591 Poquonnock Road	Hackensack	NJ	64106	Air Correction Division, UOP Inc.
Slocum, Jr.	Ernest F.	433 Hackensack Avenue	Kansas City	MO	60018	Black & Veatch Constr. Engr.
Smith	Craig	324 E. 11th Street	Des Moines	IA	64114	Ecolaire Canada Ltd.
Smith	Dan B.	999 Touhy Avenue	Kansas City	MO	CANADA	American Air Filter
Smith	Earl O.	1500 Meadowlake	Mississauga	Ontario	40277	
Smith	G. H.	33 City Center Drive #358	Louisville	KY		
Smith	Gordon L.	1930 Bishop Lane				

Smith	Smith	Mark D.	421 South 500 East	Salt Lake City	UT	84102	Brock, Easley, Inc.
Smith	Norman B.	Peter V.	Stanley Building	Muscataine	IA	52761	Stanley Consultants Inc.
Smith	Phil	Roger	P. O. Box 1500	Somerville	NJ	08876	Research Cottrell
Smith	Russell K.	Scott	525 South Hayden Road	Tempe	AZ	85281	Fabric Filters
Smith	Sidney T.	G. Ray	505 W. King	Columbus	OH	43201	Battelle Columbus Laboratories
Smith	A. J.	R. Bruce	P. O. Box 880	Addison	TX	75001	Metal Components
Smithson, Jr.	George A.	Ajay	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell
Snyder	James P.	Larry J.	1000 Prospect Hill Road	Columbus	OH	43201	Battelle Columbus Laboratories
Sommer	Stan	Harold C.	121 SW Salmon	Windsor	CT	06095	Combustion Engineering, Inc.
Sood	William D.	John	1800 FMC Drive, West	Portland	OR	97204	Portland General Electric Company
Spellman	John	F. H.	P. O. Drawer 2038	Itasca	IL	60143	FMC Corporation
Sperry	Robert M.	H. D.	Ten WOP Plaza	Pittsburgh	PA	15230	Gulf Science & Technology Company
Stachura	Peter	Edward W.	1029 Corporation Way	Des Plaines	IL	60016	Air Correction Div., UOP Inc.
Stalter	Richard D.	Nicholas J.	8900 De Soto	Palo Alto	CA	94303	Research-Cottrell, Inc.
Stanbro	Dorothy A.	Gerald W.	P. O. Box 932	Canoga Park	CA	91304	Rockwell-Energy Systems Group
Stange	Robert E.	Norman D.	Appl. Physics Lab., Johns Hopkins Rd	Laurel	MD	20810	Hiles Laboratories, Inc.
Statnick	Joseph P.	Jerome	393 7th Avenue	New York	NY	10001	Gibbs & Hill Inc.
Steeves	Harry A.	Erwin R.	50 Beale Street	San Francisco	CA	94119	Rechtel Power Corporation
Steiner	Joseph J.	Robert L.	401 H Street, S. W.	Washington	DC	20460	U.S. EPA - OEH
Stenby	Robert L.	Y. P.	P. O. Box 299	Hahone Bay	NJ	07039	Atlantic Bridge Company Limited
Sterngle	Kim	Russell L.	12 Beach Tree Road	Livingston	CO	80217	Foster Wheeler Dev. Corp.
Stern	Donald O.	Donald E.	One Penn Plaza	Denver	CO	80217	Stearns-Roger Inc.
Stevens	Kim	Donald E.	MD-61	New York	NY	10001	Chemico Air Pollution Control
Stewart	Donald E.	John C.	3412 Hillview	Res. Tri. Park	NC	27711	U.S. EPA, IERI-RTP
Stewart	Robert E.	L. A.	P. O. Box 880	Somerville	NJ	08876	Research-Cottrell
Stout	Robert E.	Louis	709 Cedar Way	Palo Alto	CA	94303	Electric Power Research Institute
Stowe	Donald H.	Joseph P.	P. O. Box 13138	Morgantown	WV	26505	Department of Energy
Strakey, Jr.	Jerome	Harry A.	4500 Cherry Creek	Oakmont	PA	15139	Chemsteel Construction Co., Inc.
Straw	Erwin R.	Joseph J.	650 Smithfield Street	Atlanta	GA	30324	Boiler Equipment
Strong	Robert L.	Y. P.	4800 Forbes Avenue	Denver	CO	80217	Dravo Lime Company
Sturtevant	Kim	Russell L.	6621 Electronic Drive	Pittsburgh	PA	15222	U.S. Department of Energy
Su	Donald O.	Donald E.	1007 Market Street	Springfield	VA	22151	Versar, Inc.
Swahlstedt	Donald E.	John C.	P. O. Box 400	Wilmington	DE	19898	E.I. du Pont de Nemours & Co., Inc.
Swartz	Robert L.	Y. P.	P. O. Box 1107, Tokeneke Road	Naperville	IL	60540	Amoco Oil Company
Swenson	Kim	Russell L.	P. O. Box 6,	Darien	CT	06820	Air Correction Division, UOP Inc.
Syler	Donald E.	John C.	300 Liberty Street	Solvay	NY	13209	Allied Chemical Corporation
Takoryan	Robert L.	Y. P.	P. O. Box 1764	Houston	TX	77001	Brown & Root, Inc.
Tanner	Robert L.	Y. P.	1500 E. Putnam	Peoria	IL	61602	Central Illinois Light Company
Tao	Robert L.	Y. P.	P. O. Box 538	Baltimore	MD	21203	Environmental Elements Corporation
Temnyson	Robert L.	Y. P.	P. O. Box 6428	Kansas City	MO	64114	Black & Veatch
Thaxton	Robert L.	Y. P.	420 Rouser Road	Jackson	MI	49201	Consumers Power Company
Theodore	Robert L.	Y. P.	Manhattan College Campus	Old Greenwich	CT	06870	Flakt Inc.
				Houston	TX	77001	Gravon Tank & Mfg. Company
				Allentown	PA	18104	Air Products & Chemicals
				Fort Myers	FL	33901	The Hunters Corporation
				Coraopolis	PA	15708	Envirotech Corporation
				Riverdale	NY	10671	Manhattan College

Thompson	Charles C.	9041 Executive Park Drive	Knoxville	TN	37919	United Engineers & Constructors
Tieman	John W.	350 Hochberg Road	Monroeville	PA	15146	Bituminous Coal Research, Inc.
Toedman	John R.	P. O. Box 652	Princeton	NJ	08540	Princeton Chemical Research, Inc.
Tokerud	Arvid	Ole Deggs 10	Oslo 6	NORWAY		Flakt Norsk Viftefabrikk A/S
Toler	Helen G.	P. O. Box 12194	Res. Tri. Park	NC	27709	Research Triangle Institute
Toor	Herbert L.	Schenley Park	Pittsburgh	PA	15213	Carnegie-Mellon University
Townsend	Joanne	1111 Bonanza	Tampa	FL	33622	Las Vegas Review-Journal
Traum	Steven R.	P. O. Box 22317	Westlake Village	CA	91361	Badger America, Inc.
Trautner	Richard P.	2625 Townsgate Road, Suite 360	Palatine	IL	60067	Environ. Res. & Technology, Inc.
Treshler	Joseph R.	139 S. Linden	Denver	CO	80201	Air Correction Division, WOP Inc.
Tsutsui	Koyo	P. O. Box 400	Boise	ID	83729	Mitsubishi International Corporation
Tuttle	John D.	11499 Chester Road	Cincinnati	OH	45246	Horrison-Knudsen Company, Inc.
Twombly	Robert	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Uehara	Shigeru	118 Tomihisa-Cho Ichigaya	Shinjuku-Ku	TOKYO		Mitsubishi Heavy Industries, Ltd.
Ulrich	Charles R.	Civil Engineering Department	Louisville	KY	40208	University of Louisville
Urbanik	Bernard	175 E. 5th Street	St. Paul	MN	55101	Burlington Northern Inc.
Vail	David L.	100 Summer Street	Boston	MA	02110	United Engineers & Constructors
Van der Bruggen	F. W.	Ulrechtseweg 310	Arnhem	NETHERLANDS		N.V. KEMA
Van Horn	Andrew J.	2118 Hilvia Street	Bekeley	CA	94704	Teknekron
Van Ness	Robert P.	P. O. Box 32010	Louisville	KY	40232	Louisville Gas & Electric Company
Vasan	Srini	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Vaughn	C. F.	P. O. Box 230	Las Vegas	NV	89151	Nevada Power Company
Veesaert	Marlin J.	2500 Drilling Service Drive	Maryland Heights	MO	63043	General Aggregate Corporation
Vogelsang, Jr.	C. W.	Engineering Department	Wilmington	DE	19898	F. I. du Pont de Nemours and Co., Inc.
Von Bergen	Hick	2701 Stoughton Road	Madison	WI	53716	Warman International Inc.
Voss	Stephn C.	6320 Augusta Drive	Springfield	VA	22150	Kulak, Voss & Co., Incorporated
Walker, Jr.	Hamilton G.	P. O. Box 1318	Baltimore	MD	21203	Environmental Elements Corp.
Walters, Jr.	Richard V.	P. O. Box 7808	Boise	ID	83729	Horrison-Knudsen Company, Inc.
Wang	Shih-Chung	1102 Q Street	Sacramento	CA	95812	California Air Resources Board
Waters	R. E.	50 Roale Street	Toronto, Ontario	CANADA		Rechtel National Inc.
Webb, Jr.	Clifford A.	700 University Avenue	Hattiesburg	MS	39401	S. Mississippi Electric Power Assoc.
Webber	Emlyn	P. O. Box 1589	Horsham	PA	19044	IU Conversion Systems, Inc.
Weber	Gerhard W.	Bldg. 3, Empire State Plaza	Albany	NY	12223	NY State Public Service Commission
Weber	Henry C.	415 E. 52nd St., Ste. 1D	New York	NY	10022	H & W Management Science
Wedig	Christopher P.	225 Franklin Street	Boston	MA	02110	Stone & Webster Engineering Corp.
Weeter	Dennis W.	73 73 Perkins Hall-UT	Knoxville	TN	37916	University of Tennessee
Weiner	Paul E.	21 West Street	New York	NY	10006	Ebasco Services Inc.
Weir, Jr.	Alexander	P. O. Box 800	Rosemond	CA	91770	Southern California Edison Company
Weiss	Kenneth N.	6601 W. Broad Street	Richmond	VA	23261	Reynolds Metals Company
Weiss	Lawrence H.	747 Third Avenue	New York	NY	10017	Chem Systems Inc.
Wells	Murray	8500 Shoal Creek Blvd., Box 9948	Austin	TX	78766	Radian Corporation
West	Brian	4014 Long Beach Blvd.	Long Beach	CA	90807	SCS Engineers Inc.
West	Stephen L.	700 West State, Div. of Environment	Boise	ID	83720	Bureau of Air Quality
White	Charles S.	555 Madison Avenue	New York	NY	10022	Combustion Equipment Associates
White	Lowell D.	3422 South 700 West	Salt Lake City	UT	84119	ASARCO Incorporated
Wick	Rodger	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Wicks	W. D.	29 S. La Salle Street	Chicago	IL	60603	Babcock & Wilcox Company
	Dale A.	Stanley Building	Muscatine	IA	52761	Stanley Consultants, Inc.

Wiedersum	George	2301 Market Street	Philadelphia	PA	19101	Philadelphia Electric Company
Wilbur	John C.	P. O. Box 101	Florham Park	NJ	07932	Exxon Research & Engineering Co.
Willett	Howard	835 Hope Street	Stanford	CT	06909	Peabody Process Systems Inc.
Williams	Angela J.	2020 Bldg., Abbott Road	Midland	MI	48640	Dow Chemical Company
Wilson	D. D.	607 East Adams Street	Springfield	IL	62701	Central Ill. Public Service Co.
Wolfe	Brian A.	P. O. Box 835	Alliance	OH	44601	Babcock & Wilcox
Woodyard	John	4014 Long Beach Boulevard	Long Beach	CA	90807	SCS Engineers
Worthington	Sidney	401 H Street, S.W.	Washington	DC	20460	U.S. EPA
Wright	Charles	525 South Hayden Road	Tempe	AZ	85281	Fabric Filters
Wunder	Pat	P. O. Box 15027	Las Vegas	NV	89114	ENSL-Las Vegas, US. EPA
Yagi	S.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Yanagioka	Hiroshi	3-13 Moriya-cho Kanagawaku	Yokohama	JAPAN	221	Chiyoda Chemical Eng. & Con.Co., Ltd.
Yavorsky	John	901 Oak Tree Road	S. Plainfield	NJ	07080	ASARCO
Yosick	Paul	1930 Bishop Lane	Louisville	KY	40277	American Air Filter
Young	Darrell T.	1515 Mineral Square, Box 11299	Salt Lake City	UT	84147	Kennecott Copper Corporation
Zaharchuk	Roman	P. O. Box 699	Pottstown	PA	19464	Firestone Tire Company
Zarchy	Andrew S.	1 River Road	Schenectady	NY	12301	General Electric
Ziegenhorn	Geroge J.	400 East Sibley Blvd.	Harvey	IL	60426	Atlantic Richfield Company
Ziminski	Richard W.	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Zuchowski	Leon A.	670 Winters Avenue	Paramus	NJ	07652	Burns and Roe, Inc.
Zuckerman	I.	Environ. Engineering Div.	Redondo Beach	CA	90178	TRW